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PCLAB Manual

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ABSTRACT

A user manual for the Process Control Laboratory (PCLAB) software is composed. The PCLAB is a MATLAB-based computer program that simulates certain chemical processes for dynamic analysis and control. The manual consists of eight chapters and two appendices. The manual is organized such that the first chapter gives and overview of the MATLAB basic functions, SIMULINK, and PCLAB. The second chapter introduces the user to the case studies which comprises of seven common industrial processes. Brief theoretical background regarding dynamic analysis and control is discussed in chapter three. Chapter four addresses the steady state disturbance sensitivity analysis. Chapter five and six cover the open-loop and closed-loop dynamic analysis, respectively. Chapter seven addresses the control of multivariable systems while chapter eight investigate advanced topics such as cascade and feed-forward control configuration.

الملخص

تم اعداد دليل المستخدم لبرنامج مختبر عمليات التحكم (بي سي لاب) الذي يعتمد على البرنامج الهندسي ماتلاب و يعنى بمحاكاة العمليات الكيميائيه لدراسة سلوكها الدينامي و التحكم فيها. يتكون الكتيب من ثمانية ابواب و ملحقان. يهتم الباب الأول يتقديم كل من برنامج ماتلاب و سميولنك و بي سي لاب. و اما الباب الثاني فيستعرض سبع عمليات تصنيع كيميائيه و التي ستسخدم كاساس للدراسات التجريبيه. و في الباب الثالث يتم تقديم الأسس النظريه و المعادلات الرياضيه الخاصه بالتحكم و ديناميكية العمليات. فيما يخص تحليل حساسية الإطضرابات في حالة الإستقرار فتناقش في الباب الرابع بينما يختص الباب الخامس و السادس بعرض الدروس الخاصه بالتحليل الدينامي للدائرة المفتوحة و المغلقه. كما يخصص الباب السابع لمعالجة عمليات التحكم متعددة المتغيرات و الباب الثامن لعمليات التحكم المتقدمه مثل الدوائر المركبه و التغذيه الأماميه.

CHAPTER 1: INTRODUCTION

1.1 OBJECTIVES

The PCLAB software is developed based on MATLAB tools and functions including SIMULINK (a graphical simulation Toolbox). Therefore, the aim of this chapter is to introduce the student to MATLAB and SIMULINK as well as explain the basic operations and installation of the PCLAB software. A brief description of MATLAB and SIMULINK will be presented, further information can be found in printed document or online documents provided by MATLAB Inc.

1.2 WHAT IS MATLAB

MATLAB is developed by Math Works Inc. It contains over 200 mathematical subprograms that enables the engineer to solve a broad range of mathematical problems including complex arithmetic, eigen-value problems, differential equations, linear and non-linear systems and many other special functions.

MATLAB is a powerful tool that is widely applied to solve many problems in process control and dynamics. In order to be able to fully apply PCLAB, one must be familiar with MATLAB commands. A few examples of the MATLAB commands are given below to familiarize the user with the software and as one practices, he becomes exposed to more and more commands.

When the software is loaded to the computer memory, the prompt '>>' is displayed. The program is in an interactive command mode. If an expression is entered with the correct syntax, it is processed immediately and the result displayed on the screen. The tutorials below are intended to introduce the reader to MATLAB so as to give the students handy experience with the features that are essential to understanding and evaluation of the software.

1.3 BASIC OPERATIONS AND COMMANDS

The four elementary arithmetic operations in MATLAB are done by the operators +, -, *, /, and ^, where, ^ , stands for power operator: For example, type the following and then press 'Enter' key and observe the result.

```
»2+3*4^(1-1/5)
```

The operator \ is for left division. For example, try:

```
»2/4
»2\4
```

Example 1. Using MATLAB as a calculator

At the prompt type 4321+8765

```
>>4321+8765 (hit enter)

Ans =  13086
```

Or try to solve the expression 2*3+4/5*1.5+11

```
>>2*3+4/5*1.5+11  (hit enter)
ans = 18.2
```

MATLAB easily handles complex and infinite numbers:

```
»sqrt( -1)
Ans = 0 + 1.0000i
```

Both *i* and *j* stand for the complex number (that is square root of –1) unless another value is assigned to them. Also the variable *pi* represents the ratio of the circumference of a circle to its diameter (i.e., 3.141592653.. .). If an expression cannot be evaluated, MATLAB returns NaN, which stands for Not-a-Number:

The equality sign is used to assign values to variables:

```
»a = 2
»b = 3*a
```

If no name is introduced, result of the expression is saved in a variable named *ans:*

```
»a+b
```

If you do not want to see the result of the command, put a semicolon at the end of it:

```
»a+b;
```

You can see the value of the variable by simply typing it:

»a

MATLAB is case sensitive. This means MATLAB distinguishes between upper and lower case variables. In MATLAB all computations are done in double precision. However, the result of calculation is normally shown with only 5 digits. The *format* command may be used to switch between different output display formats:

»c = *exp(pi)*

»format long, c

»format short e, c

»format long e, c

»format short, c

The *clc* command clears the command window and homes the cursor:

»clc

Remember that by using the up arrow key you can see the commands you have entered so far in each session. If you need to call a certain command that has been used already, just type its first letter (or first letters) and then use the up arrow key to call that command.

Several navigational commands from DOS and UNIX may be executed from the MATLAB Command Window, such as *cd, dir, mkdir, pwd, Is.* For example:

»cd d:\matlab\toolbox

»cd 'c:\Program Files\Numerical Methods\Chapterl '

The single quotation mark (') is needed in the last command because of the presence of blank spaces in the name of the directory.

1.3.1 Vectors and Matrices

MATLAB is designed to make operations on matrices as easy as possible. Most of the variables in MATLAB are considered as matrices. A scalar number is a 1 x 1 matrix and a vector is a 1 x n (or n x 1) matrix. Introducing a matrix is also done by an equality sign:

```
»m = [1 2 3; 4, 5, 6]
```

Note that elements of a row may be separated either by a space or a comma, and the rows may be separated by a semicolon or carriage return (i.e., a new line). Elements of a matrix can be called or replaced individually:

```
»m(1,3)
»m(2,1) = 7
```

Matrices may combine together to form new matrices:

```
»n = [m; m]
»o = [n, n]
```

The transpose of a matrix results from interchanging its rows and columns. This can be done by putting a single quote after a matrix:

```
»m = [m; 7, 8, 9]'
```

A very useful syntax in MATLAB is the colon operator that produces a row vector:

```
»v = -1:4
```

The default increment is 1, but the user can change it if required:

```
»w = [-1:0.5:4; 8:-1:-2; 1:11]
```

A very common use of the colon notation is to refer to rows, columns, or a part of the matrix:

```
»w(:,5)
»w(1,:)
»w(2:3,4:7)
```

```
»w(2,8:end)
```

Two useful array functions are *size,* which gives the size of the array, and *length,* which gives the maximum length of the array:

```
»size(w)
»length(w)
```

1.3.2 Array Arithmetic

Multiplying a scalar to an array, multiplies all the elements by the scalar:

```
»a = [1, 2, 4; 2:4; 4:0.5:5], 2*a
```

Only two arrays of the same size may be added or subtracted:

```
»b = ones(3); a-b
»c = ones(3,1); a-c
```

Adding a scalar to an array results in adding the scalar to all the elements of the array:

```
»a+2
```

Vector and matrix multiplication requires that the sizes match:

```
»a*b
»c'*a
```

To perform an operation on an array element-by-element, use a "." before the operator:

```
»a.*b
»a^2
»a.^2
» l./a
```

Some useful matrix functions are:

»det(a)	determinant of a square matrix
»inv(a)	inverse of matrix
»rank(a)	rank of matrix
»eig(a)	eigenvalues and eigenvectors of square matrix

»poly(a) characteristic polynomial of matrix

»svd(a) singular value decomposition

Example 2. Using MATLAB for matrix calculations

Enter the example below at the prompt

```
>> b = [1 2;3 4] (hit enter)
b =  1  2
      3  4
c = [5 6; 7 8] ( hit enter)
c =  5  6
     7  8
>> d = b*c  (hit enter)
ans = 19  22
        43 50
```

To find the transpose of d, type d'

```
>> d'  (enter)
ans =  19   43
         22   50
```

Example 3 Using MATLAB elementary math functions

The software also has several elementary math functions. These include trigonometric functions, exponential functions and other built-in matrix functions like determinant, rank and null-space. Examples of trigonometric and exponential functions are:

Trigonometric functions

```
>> cos(2)  (hit enter)

ans= -0.4161
```

>>sin(pi) (hit enter)

ans = 1.2246e-16

Exponential functions

>>exp(1) (hit enter)
ans = 2.7183
>>log(2) (hit enter)
ans = 0.6931

1.3.3 Graphics

The program also has outstanding graphic capabilities. It can be done with automatic scaling or within defined scales as shown in the example below

>> x = -10: .1 :10;
>>plot (x.^2)
>>figure
>>plot (x, x.^2)
>>figure

1.3.3.1 Plotting graphs

2-D Graphs: Functions with one independent variable can easily be visualized in MATLAB:

»x = *linspace*(0,2)*;* y = *x.*exp*(*-x*)*;*

»plot(y)	plots *y* versus their index
»plot(x,y)	plots *y* against *x*
»grid	adds grid lines to the current axes
»grid	removes the grid lines
»xlabel('x')	adds text below the x-axis
»ylabel('y')	adds text besides the y-axis
»title('y = x*exp(-x)')	adds text at the top of the graph
»gtext('anywhere')	places text with mouse

»text(1,0.2,’(1,0.2)’) places text at the specific point

You can use symbols instead of lines. You can also plot more than one function in a graph:

»plot(x,y,'.' ,x, x.*sin(x))

Also, more than one graph can be shown in different frames:

*»subplot(2,1,1), plot(x,x.*cos(x))*

*»subplot(2,1,2), plot(x,x.*sin(x))*

Axis limits can be seen and modified using the *axis* command:

»axis

»axis([0*,* 1.5,0, 1.5])

Before continuing, clear the graphic window:

»clf

Another easy way to plot a function is:

*»fplot('x*exp(-x)',[0,2])*

The function to be plotted may also be a user-defined function.

1.3.4 Scripts and Functions

The programs written in the language of MATLAB should be saved with the extension of *"m",* from where the name of *m-file* comes. When using the editor of MATLAB, it automatically saves your files with the *"m"* extension. Otherwise, one should be sure to save them with the *"m"* extension. M-files can be in the form of *scripts* and *functions* and could be executed in the MATLAB workspace.

A script is simply a series of MATLAB commands that could have been entered in the workspace. When typing the name of the script, the commands will be executed in their sequential order as if they were individually typed in the workspace. For example, let us calculate the volume of an ideal gas as a function of pressure and temperature. Type the following commands in the editor and save it as *"myscript.m":*

```
% A sample script file
disp(' Calculating the volume of an ideal gas.')
R = 8314;    % Gas constant
t = input(' Vector of temperatures (K) = ');
p = input(' Pressure (bar) = ')*le5;
v = R *t/p;        % Ideal gas law
% Plotting the results plot(t,v)
xlabel('T (K)')
ylabel('V (m^3/kmol)')
title('Ideal gas volume vs temperature')
```

The symbol % indicates that this line contains comments. The % sign, and what comes after it in that line, will be ignored at the time of execution. Return to the MATLAB command window and type:

»myscript

Input the required data and see the results.

1.3.5 Flow Control

MATLAB has several flow control structures that allow the program to make decisions or control its execution sequence. These structures are *for, if, while,* and *switch* which we describe briefly below:

If....(else...) end -The *if* command enables the program to make decision about what commands to execute:

```
x=input(' x= ');
if x >= 0
y = x^2
end
```

You can also define an *else* clause, which is executed if the condition in the *if* statement is not true:

```
x = input(' x = ');

if x>=0

y = x^2

else
y = -x^2
end
```

for….end –The for command allows the script to cause a command, or a series of commands, to be executed several times:

```
k=0;
for x = 0:0.2:1
k=k+l
y(k) = exp( -x)
end
```

while…end -The while statement causes the program to execute a group of commands until some condition is no longer true:

```
x=0;
while x<l
y= sin(x)
x = x+0.1;
end
```

switch…case….end -When a variable may have several values and the program has to execute different commands based on different values of the variable, a *switch-case* structure is easier to use than a nested *if* structure:

```
a = input('a = '); switch a case 1
disp('One'); case 2
disp('Two')
```

```
case 3
disp('Three')
end
```

Two useful commands in programming are *break* and *pause.* You can use the *break* command to jump out of a loop before it is completed. The *pause* command will cause the program to wait for a key to be pressed before continuing:

```
k=0;
for x = 0:0.2:1
if k>3
break
end
k = k+l
y(k) = exp( -x)
pause
end
```

1.3.6 Data Export and Import

There are different ways you can save your data in MATLAB. Let us first generate some data:

```
»a = magic(3); b = magic(4);
```

The following command saves all the variables in the MATLAB workspace in the file *"f*1.*mat':*

»save f1

If you need to save just some of the variables, list their names after the file name. The following saves only *a* in the file *"f2.mat":*

```
»save f2 a
```

The files generated above have the extension *".mat"* and could be retrieved only by MATLAB.

To use your data elsewhere you may want to save your data as text:

»save f3 b *-ascii*

Here, the file *"f* 3" is a text file with no extension.

You can load your data into the MATLAB workspace using the *Load* command. If the file to be loaded is generated by MATLAB (carrying *".mat"* extension), the variables will appear in the workspace with their name at the time they were saved:

»clear

»load fl

1.4 INTRODUCTION TO SIMULINK

SIMULINK is an interactive programming environment within MATLAB which provides graphical interface to MATLAB programs. The program is user-friendly and one can run simulations of linear and nonlinear systems interactively. The system can then be analyzed and the results easily visualized by simple click-and-drag of mouse operations.

Although the standard MATLAB package is useful for analysis of linear systems, SIMULINK is far more useful for the simulation of control system. It enables the rapid construction and simulation of control block diagrams. In order to make it easy for learning SIMULINK, the reader is taken through different tutorials. The goal of the tutorial is to introduce the use of SIMULINK for control system simulation. The tutorials are divided into sections namely; Background, Open-loop Simulations, Closed-loop Simulations

1.4.1 Background

The first step is to startup MATLAB on the machine you are using. In the Launch Pad window of the MATLAB desktop, select SIMULINK and then the SIMULINK Library Browser. This is done through the steps below:

1. In your PC, double click the MATLAB icon. As a result, a MATLAB command window pops up and becomes ready for operation.

2. Change the operating directory to the subdirectory under which a specific tutorial exist by typing, At the MATLAB command prompt, the following:

   >> cd C:\workshop\simulink\MPC
   (This will put you in the subdirectory designated for Model predictive simulations)

   Or
   >> cd C:\workshop\simulink\ident

(This will put you in the subdirectory designated for identification simulations)

Or

>> cd C:\workshop\simulink\phlink

(This will put you in the subdirectory designated for pH process simulations)

etc…

While in the SIMULINK Library Browser number of options are listed, as shown in Figure 1.1. Notice in this figure that **Continuous** has been highlighted; this will provide a list of continuous function blocks available. Selecting Continuous will provide the list of blocks shown in Figure 1.2. The ones that we often use are **Transfer Fcn** and **State-Space**.

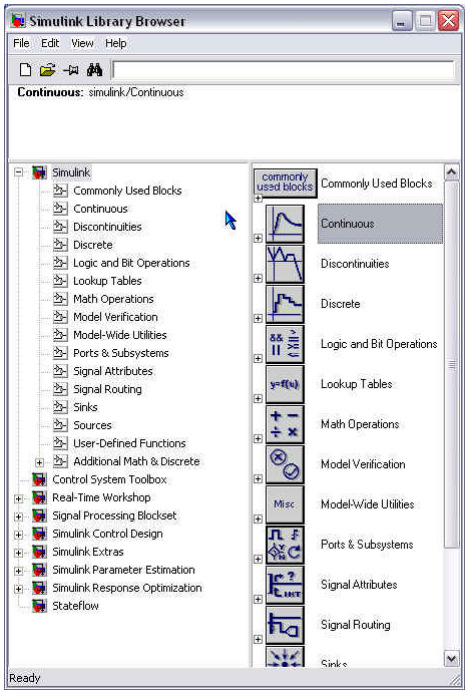


Fig. 1.1 SIMULINK Library Browser

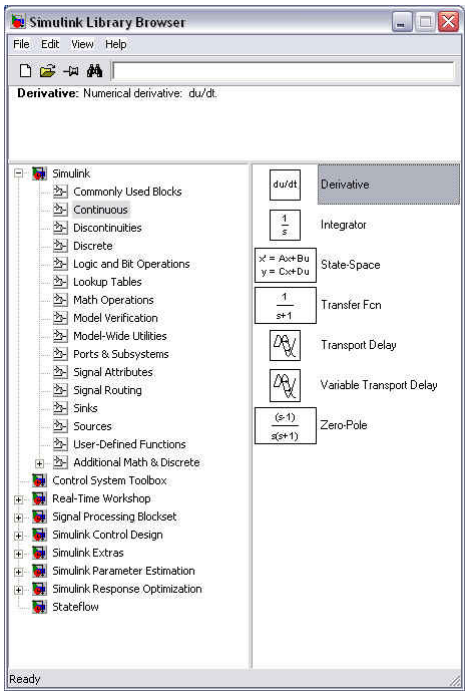


Fig. 1.2 SIMULINK Continuous Blocks.

Selecting the **Sources** icon yields the library shown in Fig. 1.3. The most commonly used sources are **Clock** (which is used to generate a time vector), and **Step** (which generates a step input).

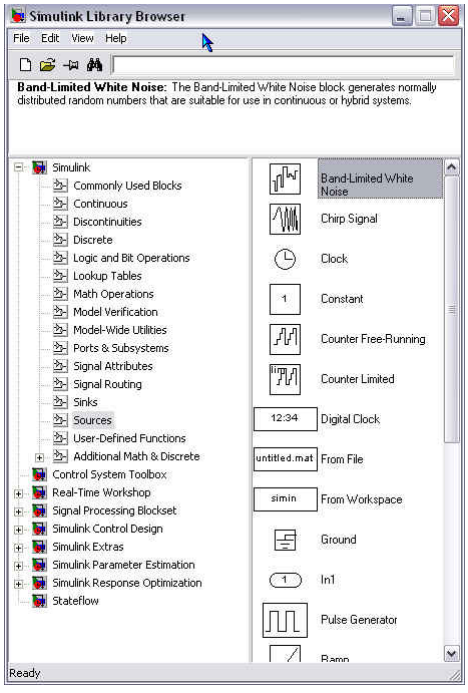


Fig. 1.3 SIMULINK Sources Library

The Sinks icon from Fig. 1.1 can be selected to reveal the set of sinks icons shown in Fig. 1.4. The one that we use most often is the To Workspace icon. A variable passed to this icon is written to a vector in the MATLAB workspace. The default data method should be changed from "**structure**" to "**matrix**" in order to save data in an appropriate form for plotting.

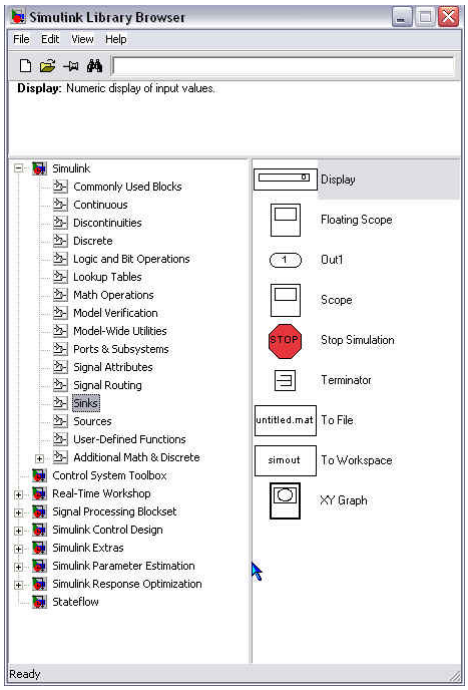
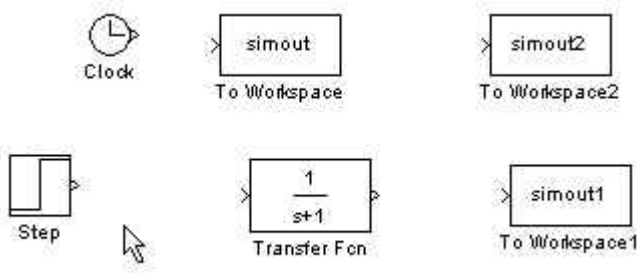


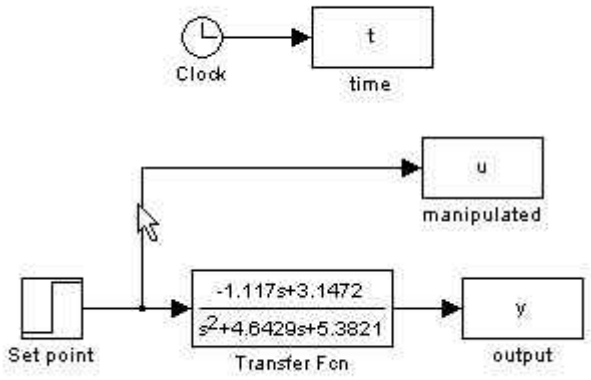
Fig. 1.4 SIMULINK Sinks Library.

1.4.2 Open-Loop Simulation

A user now has enough information to generate an open-loop simulation. The **Clock, simout, step and Transfer function** blocks can be dragged to a model **(.mdl)** workspace, as shown in Fig. 1.5a. Renaming the blocks and variables, and connecting the blocks, produces the model shown in Fig. 1.5b.



a. Placement of function blocks



b. Renaming and connecting of blocks

Fig. 1.5 Development of an Open-loop Simulation

The s-polynomials in the process transfer function were entered by double-clicking on the **transfer function** icon and entering the coefficients for the numerator and denominator of the polynomials. Notice also that the default step (used for the step input change) is to step from a value of 0 to a value of 1 at t = 1. These default values can be changed by double-clicking the **step icon**. The simulation parameters can be changed by going to the Simulation "pull-down" menu and modifying the **stop time** (default = 10) or the **integration solver method** (default = ode45).

The reader should generate simulations and observe the "inverse response" behavior of the output with respect to a step input change. Use the **subplot** command to place the process output (y) on the top plot, and the manipulated input (u) on the bottom plot. If desired, change the default simulation stop time by selecting the parameters "pull down" menu.

1.4.3 Feedback Control Simulation

The Math icon from Fig. 1.2 can be selected, resulting in the functions shown in Fig. 1.6. Additional icons can be found by selecting the **Simulink Extras icon** in Fig. 1.1. Selecting the **Additional Linear icon** from this group yields the set of icons in Fig. 1.7. The most useful icon here is the PID Controller. Any icon can be "dragged" into the

untitled model workspace. In Fig. 1.8 we show the preliminary stage of the construction of a control block diagram, where icons have been dragged from their respective libraries into the **untitled** model workspace.

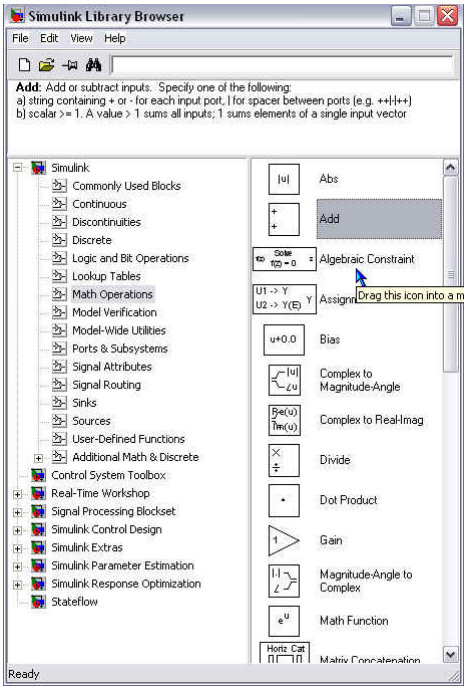


Fig. 1.6 SIMULINK Math Library.

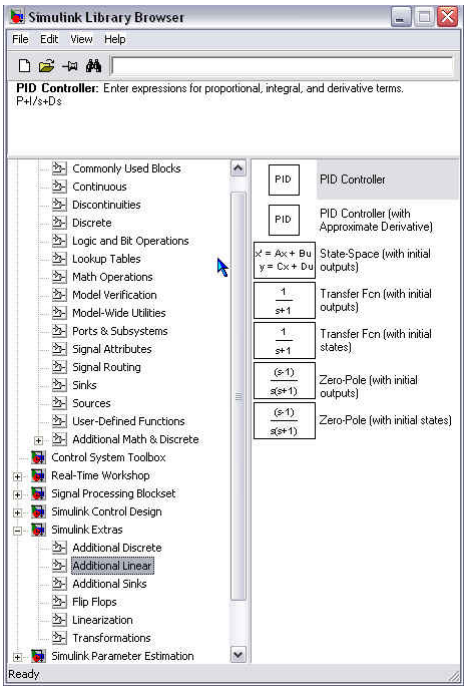


Fig. 1.7 SIMULINK Additional Linear Library.

The labels (names below each icon) can easily be changed. The default parameters for each icon are changed by double clicking the icon and entering new parameter values. Also, connections can be made between the outputs of one icon and inputs of another. Fig. 1.8b shows how the icons from Fig. 1.8a have been changed and linked together to form a feedback control block diagram.

It should be noted that the form of the PID control law used by the SIMULINK **PID Controller** icon is not the typical form that we use as process control engineers. The form can be found by double-clicking the icon to reveal the following controller transfer function representation while we normally deal with the following PID structure $P = k_c$, $I = k_{cI}$, $D = k_{cD}$.

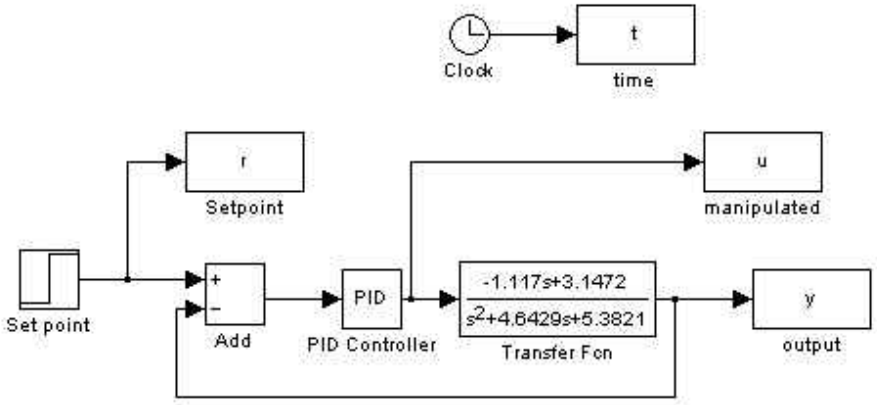


Fig. 1.8. Block Diagram for Feedback Control of the Van de Vusse CSTR

The s-polynomials in the process transfer function were entered by double-clicking on the transfer function and entering the coefficients for the numerator and denominator polynomials.

Notice also that the default step (used for the step set-point change) is to step from a value of 0 to a value of 1 at t = 1. These default values can be changed by double-clicking the step icon. The simulation parameters can be changed by going to the Simulation "pull-down" menu and modifying the stop time (default = 10) or the integration solver method (default = ode45).

Double click the PID controller box and enter the controller tuning parameters of k_c = 1.89 and τ_I = 1.23 by entering P = 1.89 and I = 1.89/1.23 in the default PID Controller block and run the simulation by clicking the start button. Generate Fig. 1.9 by typing the following on the command window.

```
» subplot(2,1,1),plot(t,r,'--',t,y)
» xlabel('t (min)')
» ylabel('y (mol/l)')
» subplot(2,1,2),plot(t,u)
» xlabel('t (min)')
» ylabel('u (min^-1)')
```

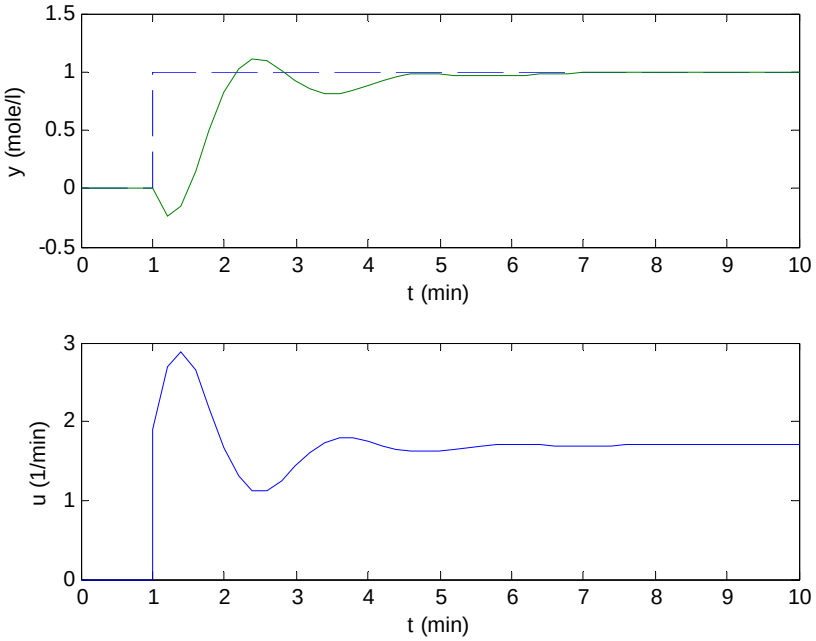


Fig. 1.9. Measured Output & Manipulated Input Responses to a Unit Set point Change.

1.4.4 Other Commonly Used Icons

Often you will want to simulate the behavior of systems that have time delays. The Transport Delay icon can be selected from the Continuous library shown in Fig. 1.2. The transport delay icon is shown in Fig. 1.10. Our experience is that simulations can become somewhat "flaky" if 0 is entered for a transport delay. We recommend that you remove the transport delay block for simulations where no time-delay is involved.

Fig. 1.10 Transport Delay Icon

Manipulated variables are often constrained to between minimum (0 flow, for example) and maximum (fully open valve) values. A saturation icon from the Nonlinear library can be used to simulate this behavior. The saturation icon is shown in Fig. 1.11.

Fig. 1.11. Saturation Element

Actuators (valves) and sensors (measurement devices) often have additional dynamic lags that can be simulated by transfer functions. These can be placed on the block diagram in the same fashion that a transfer function was used to represent the process earlier.

It should be noted that icons can be "flipped" or "rotated" by selecting the icon and going to the format "pull-down" menu and selecting Flip Block or Rotate Block. The block diagram of Fig. 1.8 has been extended to include the saturation element and transport delay, as shown in Fig. 1.12.

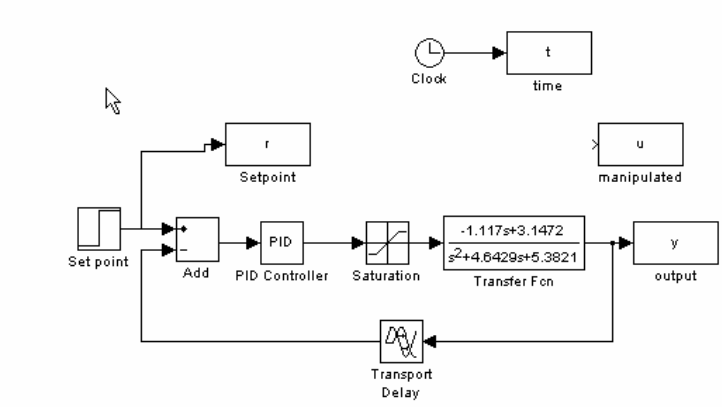


Fig. 1.12. Block Diagram with Saturation and Time-Delay Elements

The default data method for the "to workspace" blocks (r,t,u,y in Fig. 1.12) must be changed from "structure" to "matrix" in order to save data in an appropriate form for plotting.

In conclusion, SIMULINK is a very powerful block diagram simulation language. Simple simulations, including the majority of those used as examples in this textbook, can be set-up rapidly (in a matter of minutes). The goal of this module was to provide enough of an introduction to get you started on the development of open- and closed-loop simulations. With experience, the development of these simulations will become second-nature. It is recommended that you perform the simulations shown in this module, as well as other exercises, to rapidly acquire these simulation skills.

1.5 INSTALLING PCLAB

For better performance, simply copy the PCLAB folder into your local hard disk. Then either add this directory to your MATLAB path using the *path* command or choose to work from the designated directory for PCLAB using the *cd* command. Note that the working directory can be adjusted directly from the menu of the MATLAB Command Window. PCLAB works for MATLAB version 5.3 and higher.

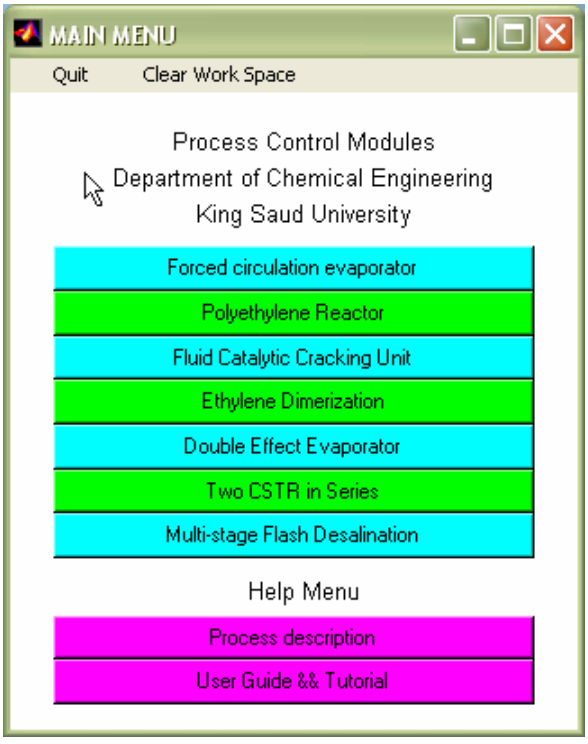


Fig. 1.13 Main menu of the PCLAB software

To start PCLAB, first lunch your MATLAB. Once in the MATLAB program, use the path browser to change to PCLAB subdirectory. Then type **mainmenu** at the MATLAB command prompt, this will take you to the itemized menu of PCLAB from which you can select the case study you want to work on. Fig. 1.13 shows the main menu of PCLAB.

1.6 OVERVIEW OF PCLAB STRUCTURE

PCLAB is interactive simulation software for process control analysis and training. The process control module contains several exercises that cover basic concept of process dynamic and control. These exercises can be carried out on simulated processes that are very common in the chemical engineering industry such as distillation columns, reactors and evaporators. The exercises are unified for all case studies. This means that the user can choose a relevant case study and work on the pre-designed exercises. PCLAB is based on menu structure that allows the user to navigate through various case studies each of which has its own submenu that comprises the designated exercises. By practicing on these tutorials, the user then learns different aspects of dynamic analysis and control design.

The structure of the software is designed in a user–friendly, menu-driven framework such that the process engineer can easily navigate through the various parts of the program, carry out simulation experiments, visualize the results and draw conclusions on the effect of different design parameters, and control configurations.

There is a Main menu (Fig. 1.13) that allows the user to choose from different case studies. The case studies covered by the current version of PCLAB are as follows:

1. Forced-Circulation Evaporator
2. Polyethylene Reactor
3. Fluid catalytic Cracking Unit
4. Ethylene Dimerization Process
5. Double Effect Evaporator
6. Two-CSTR in Series
7. Multi-Stage Flash Desalination Plant

These basic case studies represents the heart of the chemical engineering industry can be used in combination with the exercises to generate a whole lot of scenarios that will enhance the understanding of the basics of control process analyses.

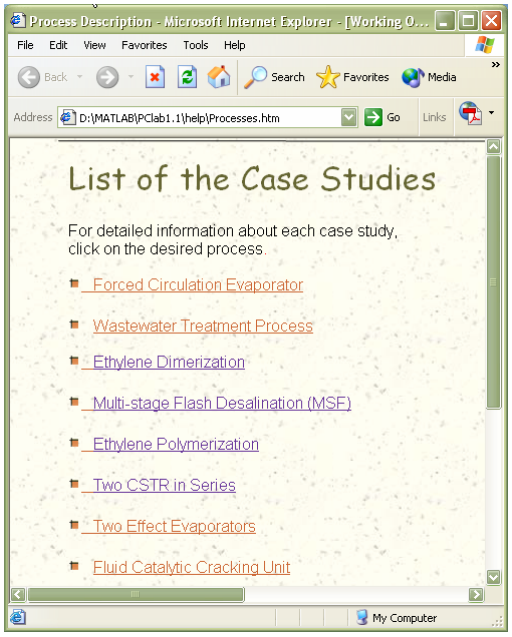

Process Description - Microsoft Internet Explorer - [Working O...
File Edit View Favorites Tools Help
Back Search Favorites Media
Address D:\MATLAB\PClab1.1\help\Processes.htm
Go Links
List of the Case Studies
For detailed information about each case study, click on the desired process.
Forced Circulation Evaporator
Wastewater Treatment Process
Ethylene Dimerization
Multi-stage Flash Desalination (MSF)
Ethylene Polymerization
Two CSTR in Series
Two Effect Evaporators
Fluid Catalytic Cracking Unit
My Computer

Fig. 1.14 Process description help option

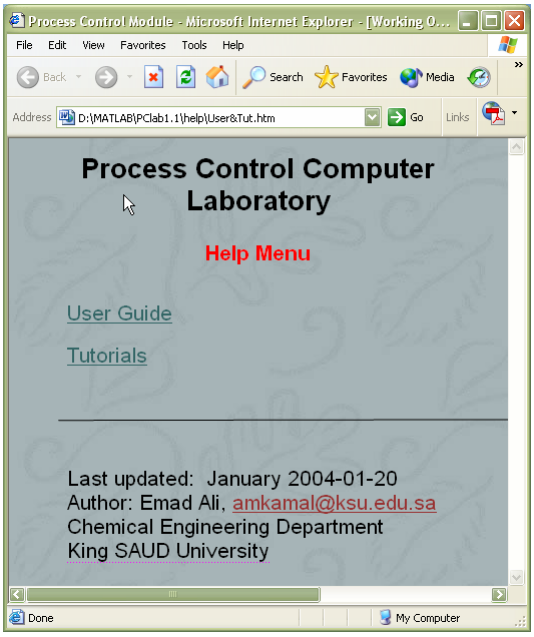


Fig. 1.15 User guide help option

The main menu has two Help option's buttons namely process description and user guide & tutorial. The help documents can be accessed online using typical web browser. The first help option (fig. 1.14) provides online information about case studies such as flow sheet, modeling equations, and typical values of the process parameters. These pieces of information can be accessed by double clicking the desired case study name. The second option (fig. 1.15) has two sub-options which provide overview of PCLAB and procedure for carrying out certain tutorials.

By clicking one of the case studies, a sub menu will appear. See for example Fig. 1.16 which shows a typical sub-menu for the forced-circulation process. This submenu allows the user to select a specific tutorial within several exercises that are ordered as follows:

1. Open-Loop
2. Closed-Loop (PID control)

3. Multi-Loop (PID control)
4. Feed Forward Control
5. Steady State Disturbance Analysis
6. System Identification.

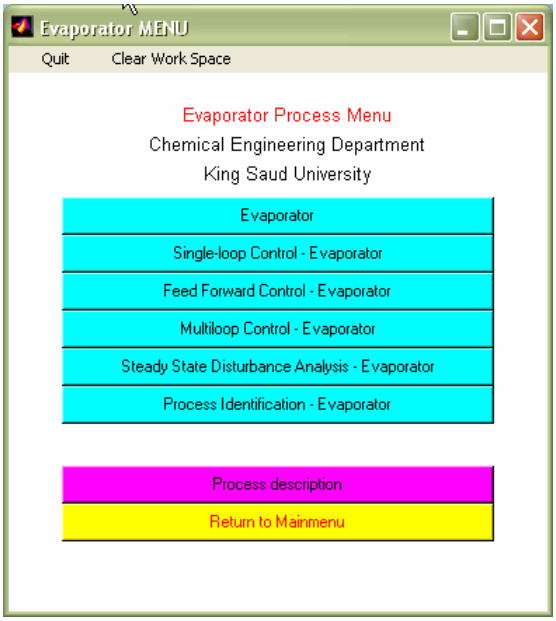


Fig. 1.16 Evaporator menu

The aforementioned exercise focuses on the following objectives:

- Modeling of the process
- Controller design and tuning
- Sensitivity analysis and system identification

The Tutorial is organized in such a way as to lead the user to the proper understanding of process dynamics. First the process is developed for steady state predictions. Since chemical process systems do not reach steady state instantaneously, accurate modeling must capture both the steady state and transient behavior of the system. The transient behavior is the time-dependent trajectory of an output in response to a particular input or class of inputs. After that control synthesis is the nest step, where the control structures are selected and tuned. Finally the designed controlled system is analyzed for proper performance. Future versions of the software can examine advance control algorithms such as Internal Model Control (IMC) as well as Model Predictive Control (MPC).

The sub menu in Fig. 1.16 has additional navigation buttons. The “process description” button allows the user to access a document file containing the description of the current active (selected) case study. The yellow button simply closes the current submenu and returns the user to the main menu.

PCLAB also includes additional features which will be discussed as we progress in the manual.

CHAPTER 2: CASE STUDIES

2.1 INTRODUCTION

In this chapter, the modeling of the basic case studies (representing the heart of the chemical engineering industry) is undertaken. Tutorials and exercises that can be carried out upon these case studies will be discussed in other chapters. These modules (case studies) are clearly defined and all involved parameters are explained. It is important to understand the mathematical development of the models as they represent the basis to which the exercises will be referred to. The mathematical models developed here for each case study are core to the process transfer functions that will be used with differing controller algorithms and are classical examples from literature.

As listed earlier, these case studies are;

8. Forced-Circulation Evaporator
9. Polyethylene Reactor
10. Waste Water Treatment Unit
11. Ethylene Dimerization Process
12. Double Effect Evaporator
13. Two-CSTR in Series
14. Multi-Stage Flash Desalination Plant

2.2 FORCED-CIRCULATION EVAPORTOR UNIT

2.2.1 Process Description and Flow sheet

The forced-circulation evaporator is a common processing unit in sugar mills, alumina production and paper manufacture. This process is used to concentrate a dilute liquor by evaporating its solvent (usually water) as shown in the Fig. 2.1.

A feed stream with solute of concentration C_1 (mass percentage) is mixed with high volumetric recycle flow rate and fed to a vertical evaporator (heat exchanger). The solution will pass through the tube. A saturated steam is used to heat up the mixture by condensing on the outer surface of the tubes. The liquor which passes up inside the tube

boils and then passes to a separator vessel. In the separator, the liquid and vapor are separated at constant temperature and pressure. The liquid is recycled with some being drawn off as product with solute concentration of C_2. The vapor is usually condensed with water. Water is used as the coolant**.**

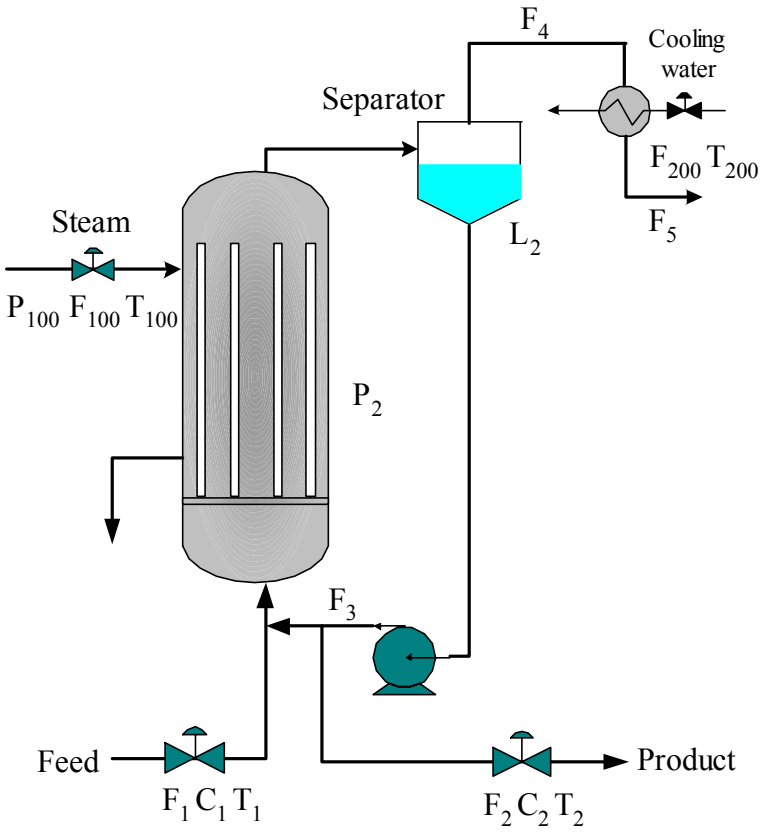


Fig. 2.1 Flow sheet of Forced Circulation Evaporator Process

2.2.2 The Process Model

$$\rho A \frac{dL_2}{dt} = F_1 - F_4 - F_2 \tag{2.1}$$

$$M \frac{dC_2}{dt} = F_1 C_1 - F_2 C_2 \tag{2.2}$$

$$W \frac{dP_2}{dt} = F_4 - F_5 \tag{2.3}$$

Additional relations:

$$T_2 = 0.5616P_2 + 0.3126C_2 + 48.43 \quad (2.4)$$

$$T_3 = 0.507P_2 + 55.0 \quad (2.5)$$

$$F_2 = F_1 - F_4 \quad (2.6)$$

$$F_4 = [Q_{100} - F_1 Cp(T_2 - T_1)]/\lambda \quad (2.7)$$

$$T_{100} = 0.1538P_{100} + 90.0 \quad (2.8)$$

$$Q_{100} = F_{100}\lambda_s = UA_1(T_{100} - T_2) \quad (2.9)$$

$$Q_{200} = F_{200}Cp(T_{201} - T_{200}) = UA_2(T_3 - 0.5*(T_{200} + T_{201})) \quad (2.10)$$

$$F_5 = Q_{200}\lambda \quad (2.11)$$

$$UA_1 = 0.16(F_1 + F_3) \quad (2.12)$$

2.2.3 Process Parameters and Variables

Variable	*Description*	*Value*	*Units*
F_1	Feed flow rate	10.0	kg/min
F_2	Product flow rate	2.0	kg/min
F_3	Circulating flow rate	50.0	kg/min
F_4	Vapor flow rate	8.0	kg/min
F_5	Condensate flow rate	8.0	kg/min
C_1	Feed composition	5.0	Percent
C_2	Product composition	25.0	Percent
T_1	Feed temperature	40.0	C
T_2	Product temperature	84.6	C

T_3	Vapor temperature	80.6	C
L_2	Separator level	1.0	m
P_2	Operating pressure	50.5	kPa
F_{100}	Steam flow rate	9.3	kg/min
T_{100}	Steam temperature	119.9	C
P_{100}	Steam pressure	194.7	kPa
Q_{100}	Heater duty	339.0	kW
F_{200}	Cooling water flow rate	208.0	Kg/min
T_{200}	Cooling water inlet temperature	25.0	C
T_{201}	Cooling water outlet temperature	46.1	C
Q_{200}	Condenser duty	307.9	kW
ρ_A	Mass holdup in the separator	20	kg/m
M	Solute holdup	20	kg
W	Constant	4	kg/kPa
λ	Latent heat of vaporization of water	38.5	kW/kg
UA_2	Heat transfer coefficient times the heat transfer area	6.84	kW/K

λ_s is the latent heat of steam which is calculated from a given correlation as a function of the saturation temperature.

2.3 ETHYLENE POLYMERIZATION REACTOR

2.3.1 Process Description and Flow sheet

Polyethylene is considered the world largest produced synthetic commodity polymer. The polyethylene reactor process is depicted in Fig. 2.2. The process model was developed by McAuley *et al*. [12] and is given below.

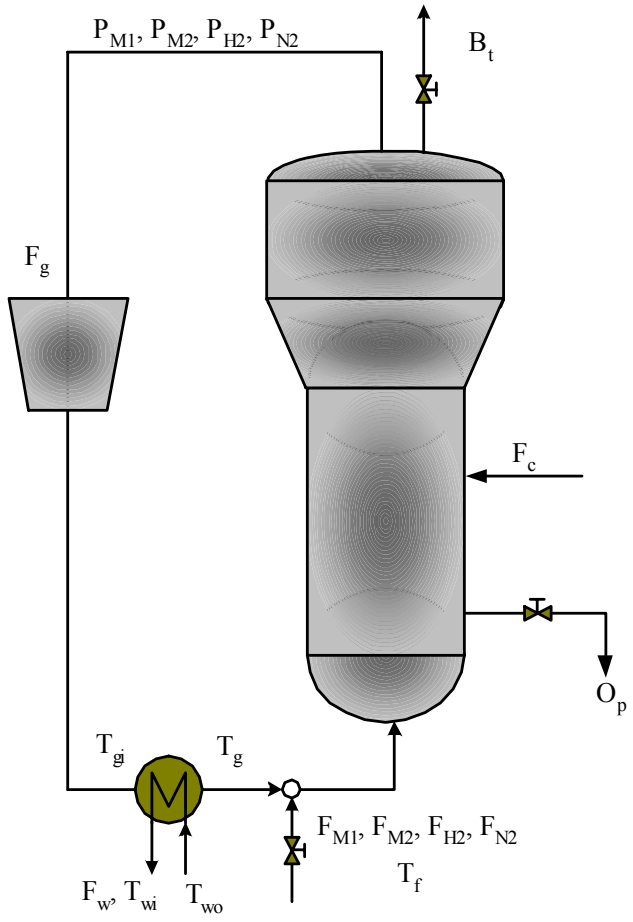


Fig. 2.2 Polymerization Reactor

The major components of the process are: (a) feed gas which is partly combined with the recycled gas before entering the bubbling fluidized bed, the other part of the fresh feed gas is used to introduce the Ziegler-Natta catalyst; (b) a catalyst feeder; (c) a product withdrawal system which is controlled in order to maintain a constant bed height inside the reactor; (d) gas recycling which includes a cyclone and a compressor; (e) reactor with catalyst disengagement zone. Four major components are fed to the reactor. The gaseous species are Ethylene (monomer), Butene (comonomer), hydrogen and nitrogen. The nitrogen is used to carry the catalyst powder and maintain the desired column pressure. In this process, the bed temperature is controlled by manipulating the feed temperature of the cooling water. The total pressure is controlled by manipulating the bleed flow. And the bed height is regulated by the discharge system. The discharge system operates periodically, i.e. the outlet flow valve opens every time the bed level exceeds certain limit.

2.3.2 The Process Model

$$V_g \frac{dC_{M1}}{dt} = F_{M1} - x_{M1}B_t - R_{M1} \tag{2.13}$$

$$V_g \frac{dC_{M2}}{dt} = F_{M2} - x_{M2}B_t - R_{M2} \tag{2.14}$$

$$V_g \frac{dC_H}{dt} = F_H - x_H B_t - R_H \tag{2.15}$$

$$V_g \frac{dC_N}{dt} = F_N - x_N B_t \tag{2.16}$$

$$\frac{dY_c}{dt} = F_c a_c - k_d Y_c - O_p Y_c / B_w \tag{2.17}$$

$$(M_r Cp_r + B_w Cp_p)\frac{dT}{dt} = HF + HG - HR - HT - HP \tag{2.18}$$

$$M_g Cp_g \frac{dT_g}{dt} = F_g Cp_g (T_{gi} - T_g) + F_w Cp_w (T_{wi} - T_{wo}) \tag{2.19}$$

$$P_t = (C_{M1} + C_{M2} + C_H + C_N)RT \tag{2.20}$$

$$T_{gi} = \left(\frac{P_t}{P_t + \Delta P}\right)^{(1-\eta)/\eta} T \tag{2.21}$$

$$F_w Cp_w (T_{wi} - T_{wo}) = 0.5UA[(T_{wo} + T_{wi}) - (T_{gi} + T_g)] \tag{2.22}$$

$$HF = (F_{M1}Cp_{M1} + F_{M2}Cp_{M2} + F_H Cp_H + F_N Cp_N)(T_f - T_{ref}) \tag{2.23}$$

$$HG = F_g Cp_g (T_g - T_{ref}) \tag{2.24}$$

$$HT = (F_g + B_t)Cp_g (T - T_{ref}) \tag{2.25}$$

$$HP = O_p Cp_p (T - T_{ref}) \tag{2.26}$$

$$HR = M_{w1} R_{M1} \Delta H_r \tag{2.27}$$

$$O_p = M_{w1}R_{M1} + M_{w2}R_{M2} \tag{2.28}$$

$$R_{M1} = C_{M1} Y_c k_{p1} e^{-\frac{E}{R}(1/T - 1/T_{ref})} \tag{2.29}$$

$$R_{M2} = C_{M2} Y_c k_{p2} e^{-\frac{E}{R}(1/T - 1/T_{ref})} \tag{2.30}$$

$$Cp_g = \sum x_i Cp_i \tag{2.31}$$

2.3.3 Process Parameters and Variables

Variable	*Description*	*Value*	*Units*
a_c	Active site concentration	0.548	mole/kg
B_w	Mass of the polymer in the bed	70.0	tonne
B_t	Bleed flow rate	10.39	mole/s
C_{M1}	Concentration of monomer,	297.06	mole/m^3
C_{M2}	Concentration of co-monomer	116.17	mole/m^3
C_N	Concentration nitrogen	166.23	mole/m^3
C_H	Concentration of hydrogen	105.78	mole/m^3
Cp_{M1}	Heat capacity of monomer	11	cal/mole K
Cp_{M2}	Heat capacity of co-monomer	24	cal/mole K
Cp_H	Heat capacity of hydrogen	7.7	cal/mole K
Cp_N	Heat capacity of nitrogen,	6.9	cal/mole K
Cp_w	Heat capacity of recycle gas and water	18.0	cal/mole K
Cp_p	Heat capacity of polymer,	0.85	cal/g K
E	Activation energy for propagation,	9000	cal/mole
F_c	Catalyst flow rate	2.0	kg/s
F_w	Cooling water flow rate,	5.6x10^5	mole/s
F_g	recycle flow rate,	8500	mole/s
F_{M1}	Monomer flow rate,	131.13	mole/s
F_{M2}	co-monomer flow rate,	3.51	mole/s
F_N	hydrogen flow rate,	1.6	mole/s
F_H	nitrogen flow rate,	2.52	mole/s
k_d	Deactivation rate constant,	0.0	1/s
k_{p1}	Propagation rate constant for monomer	85.0	L/mole s
k_{p2}	Propagation rate constant for co-monomer,	3.0	L/mole s
M_w	Water holdup in the heat exchanger,	2x10^6	mole
$M_r Cp_r$	Thermal capacitance of the reaction vessel,	14x10^6	kcal/K
Op	Polymer outlet rate,	3.6434	kg/s
P_t	Total pressure,	20.0	atm
P_{M1}	Partial pressure of monomer	8.67	atm

P_{M2}	Partial pressure of co-monomer	3.39	atm	
P_N	Partial pressure of nitrogen	4.8525	atm	
P_H	Partial pressure of hydrogen,	3.0875	atm	
R	Ideal Gas constant,	82.6×10^{-6}	atm mole	m^3/K
T	Bed temperature,	82	^oC	
T_f	feed temperature,	25	^oC	
T_{ref}	reference temperature,	87	^oC	
T_{gi}	Temperature of recycle stream before cooling	136	^oC	
T_g	Temperature of recycle stream after cooling,	51.7	^oC	
T_{wi}	Cooling water temperature before cooling,	20	^oC	
T_{wo}	Cooling water temperature after cooling,	35	^oC	
Y_c	Number of moles of catalyst site, mole	5.849	mole	
ΔH_r	Heat of reaction	-894	cal/g	

The system has two internal (built-in) PI control loops which are set as:

Control loop	k_c	τ_I
Pt → Bt	10	10
T → Tw	3	20

2.4 ETHYLENE DIMERIZATION REACTOR

2.4.1 Process Description and Flow Sheet

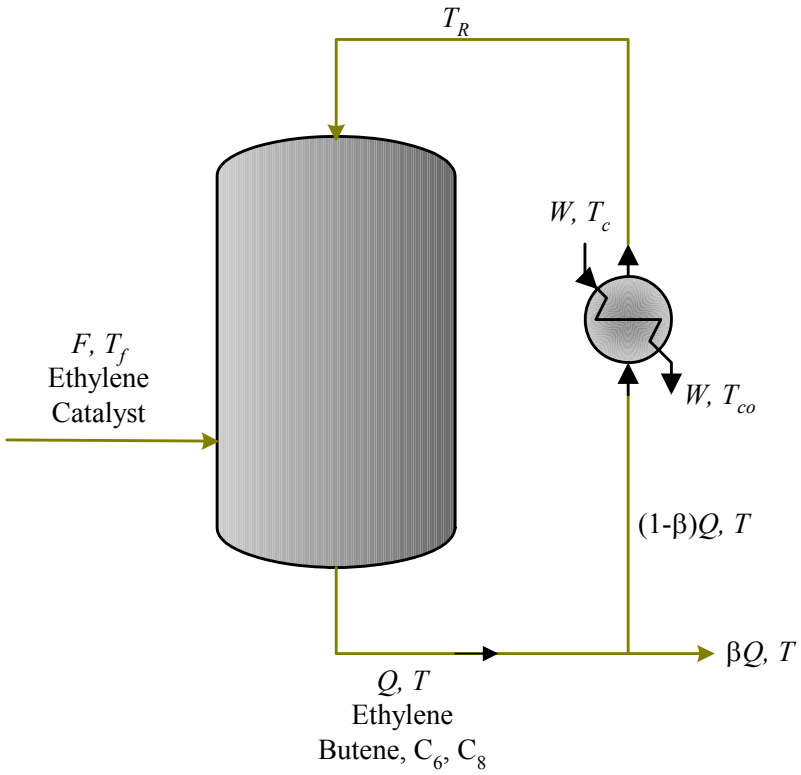


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Fig. 2.3 Dimerization Reactor

The catalytic dimerization of ethylene is considered to be one of the most promising methods for producing butene-1, the first member of the even-numbered linear 1-alkenes which have diversified applications. This process uses a homogeneous titanium-based catalyst which demonstrates high dimerization activity coupled with excellent selectivity to butene-1 at moderate pressure (20-30 psia) and temperature (50-60 ^{o}C). The dimerization reaction is regarded as a degenerate ethylene polymerization reaction and therefore the formation of heavier oligomers is expected. The industrial ethylene diemrization reactor is operating in liquid phase at bubble point conditions. Fresh ethylene and homogenous catalyst are fed continuously to the reactor where the exothermic reaction is removed by means of external loop equipped with a cooler.

The dimerization reactor considered in this study is assumed to be a liquid phase perfectly mixed reactor, i.e. no mass transfer limitation is considered in this system. Schematic of the process is depicted in Fig. 2.3. The liquid is homogenized by a high re-

circulation rate around the reactor through a heat exchange used to remove the high exothermic heat of reaction. The model uses the Homo- and Co-polymerization mechanisms suggested by Galtier *et. al*. [9]. The reaction stages; initiation, propagation and termination are of first order kinetics with respect to each reactant.

Initiation and propagation stages:

$$K + C_{2n} \rightarrow KC_{2n}, \qquad n \geq 1 \tag{2.32}$$

$$KC_{2m} + C_{2n} \rightarrow KC_{2(n+m)} \qquad n, m \geq 1 \tag{2.33}$$

The rate of initiation and propagation, which also represent the rate of disappearance of C_{2n} in these stages, have the following rate equation:

$$r_{a_{2n}} = a_{2n}[C_{2n}]\sum_{m=0}^{\infty}[KC_{2m}] \tag{2.34}$$

Termination stage:

$$KC_{2m} + C_{2n} \rightarrow K + C_{2(n+m)}, \quad n, m \geq 1 \tag{2.35}$$

The chain termination reactions, which are assumed to occur in parallel with the growth reactions have the following rate equations:

$$r_{b_{2n}} = b_{2n}[C_{2n}]\sum_{m=1}^{\infty}[KC_{2m}] \tag{2.36}$$

The rate constants are defined as follows:

$$a_2 = a_{20}e^{-E_1(\frac{1}{T}-\frac{1}{T_r})} \tag{2.37}$$

$$a_4 = a_{40} e^{-E_1(\frac{1}{T}-\frac{1}{T_r})} \tag{2.38}$$

$$b_4 = a_{40} e^{-E_1(\frac{1}{T}-\frac{1}{T_r})} \tag{2.39}$$

$$b_2 = b_{20} a_2 e^{-E_2(\frac{1}{T}-\frac{1}{T_r})} \tag{2.40}$$

Based on the above assumptions and the assumed reaction kinetics, the resulted dynamic model of the dimerization process is presented in the following [1],[6].

2.4.2 The Process Model

$$V\frac{dC_4}{dt} = -Q\beta C_4 - V[(-b_2C_2 + a_4C_4 + b_4C_4)K_2 + a_4C_4K + b_4C_4K_4] \tag{2.41}$$

$$V\frac{dC_2}{dt} = FC_{2f} - Q\beta C_2 - V[a_2C_2(K + K_2 + K_4) + b_2C_2(K_2 + K_4 + K_6)] \tag{2.42}$$

$$V\frac{dC_6}{dt} = -FC_6 + V[b_4C_4K_2 + b_2C_2K_4] \tag{2.43}$$

$$V\frac{dC_8}{dt} = -FC_8 + V[b_4C_4K_4 + b_2C_2K_6] \tag{2.44}$$

$$V\frac{dK}{dt} = FK_f - Q\beta K - V[(a_2C_2 + a_4C_4)K + b_2C_2(K_2 + K_4 + K_6) + b_4C_4(K_2 + K_4)] \tag{2.45}$$

$$V\rho C_p\frac{dT}{dt} = F\rho_f C_{pf}(T_f - T_r) + Q(1-\beta)\rho C_p(T_R - T_r) - Q\rho C_p(T - T_r) + V(r_2(-\Delta H_1) + r_4(-\Delta H_2)) \tag{2.46}$$

$$V_c\rho C_p\frac{dT_R}{dt} = Q(1-\beta)\rho C_p(T - T_R) - UA(T_{Rav} - T_{cav}) \tag{2.47}$$

$$V\frac{dK_2}{dt} = -Q\beta K_2 + V[a_2C_2K - (a_2C_2 + a_4C_4 + b_2C_2 + b_4C_4)K_2] \tag{2.48}$$

$$V\frac{dK_4}{dt} = -Q\beta K_4 + V[a_2C_2K_2 - (a_2C_2 + b_2C_2 + b_4C_4)K_4 + a_4C_4K] \tag{2.49}$$

$$V\frac{dK_6}{dt} = -Q\beta K_6 + V[a_2C_2K_4 - b_2C_2K_6 + a_4C_4K_2)] \tag{2.50}$$

where

$$T_{Rav} = \frac{T + T_R}{2} \tag{2.51}$$

$$T_{cav} = \frac{T_c + T_{co}}{2} \tag{2.52}$$

The dynamic of the outlet temperature of the coolant fluid is not included and alternatively it is obtained by solving the steady-state equation:

$$WCp_w(T_{co} - T_c) = U_h A_c(T_{Rav} - T_{cav}) \tag{2.53}$$

2.4.3 Process Parameters and Variables

Variable	*Description*	*Value*	*Units*
a_2	Rate constant for consumption of component C_2 in the initiation and propagation		(m^3/*mole s*).
a_4	Rate constant for consumption of component C_4 in the initiation and propagation		(m^3/*mole s*).
b_2	Rate constant for consumption of component C_2 in the termination stage		(m^3/*mole s*).
b_4	Rate constant for consumption of component C_4 in the termination stage		(m^3/*mole s*).
Cp	Heat capacity of reactor mixture	0.55	(*cal/gm,* ^oC)
Cp_f	Heat capacity of feed	0.55	(*cal/gm,* ^oC)
Cp_w	Heat capacity of water	1.0	(*cal/gm,* ^oC)
C_4	Butene-1 concentrations,	8700	(*mole/*m^3)
C_2	ethylene concentrations,	1065	(*mole/*m^3)
C_{2f}	Ethylene concentration at the feed	25000	(*mole/*m^3)
E_1	Activation energy for initiation & propagation reactions	-6000	K
E_2	Activation energy for the termination reaction	3000	K
F	Reactor fresh feed	0.004	(m^3/s)
K	Catalyst	1.122	(mole/m^3)
K_2	Catalyst activator for C_2	0.1345	(mole/m^3)
K_4	catalyst activator for C_4	0.0178	(mole/m^3)
K_6	Catalyst activator for C_6	0.0028	(mole/m^3)

K_f	Catalyst concentration at the fresh feed	1.25	(mole/m^3)
Q	Product flow rate	F/β	(m^3/s)
T_c	Inlet coolant temperature	0.0	(^oC)
T_f	Feed temperature	30.0	(^oC)
T_r	Reference temperature	25.0	(^oC)
T	Reactor temperature	67.0	(^oC)
T_R	Recycle temperature	43.0	(^oC)
U_hA_c	Heat transfer coefficient times heat transfer area	27500	($cal\ s\ ^oC$)
V	Reactor volume (m^3)	500	m^3
V_c	Cooler volume	50	m^3
W	Coolant flow rate,	500	(kg/s)
ρ	Mixture density	500	Kg/m^3
□ΔH_{ri}	Heat of reaction	25000	Cal/mole
β	Recycle ratio	0.02	

2.5 DOUBLE EFFECT EVAPORATOR

2.5.1 Process Description and Flow Sheet

A schematic flow sheet of the process is shown in Fig. 2.4. A solution of triethylene glycol in water is fed to the first effect at flow rate F, solute concentration C_f, and temperature T_f. The solution s concentrated in the first effect using steam at flow rate W_s, generating the overhead vapor stream O_1 and the concentrated bottom stream B_1 with solute concentration C_1. The bottom stream B_1 is fed to the tube side of the second

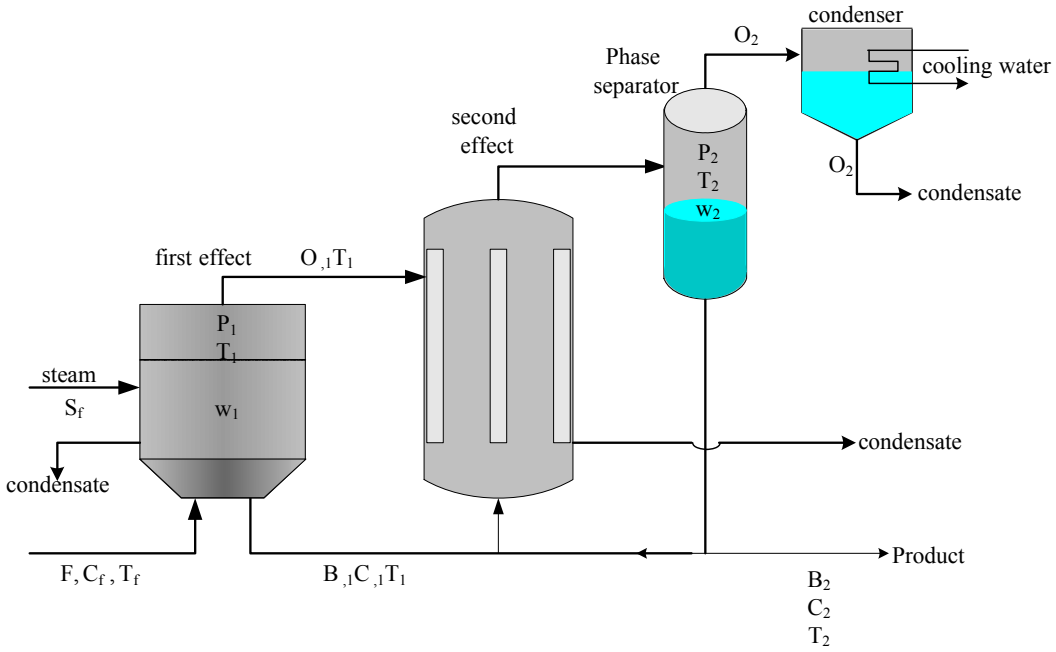


Fig. 2.4 Two Effect Evaporator

effect at a lower pressure, while the overhead stream O_1 is fed to the shell side. The bottom stream B_2, which is the product stream, leaves the second effect with solute concentration C_2. The overhead stream O_2 from the second effect is condensed and released as condensate. W_1 and W_2 are the liquid holdups, whereas P_1, T_1 and P_2, T_2 are the pressures and temperatures in the first and second effects, respectively.

A standard modeling procedure was followed to develop a dynamic model for the process using the following assumptions:

1. The steam chests, tune walls and so on have negligible heat capacity.
2. The temperature T_2 is held constant by pressure controller.

3. The overhead vapor streams leaving each of the effects have negligible solute concentration compared to the respective bottom liquid streams.
4. Vapor holdup in each effect is negligible.

Here, assumption 2 is relaxed. The stream O_2 is fixed and the temperature T_2 is evaluated such that it respects the algebraic energy balance on the second effect. In the original model, O_1 is determined from the energy balance on the first effect. Here, O_1 is determined from the pressure difference between the two effects. In addition, the bottom streams B_1 and B_2 are used to control the liquid holdups W_1 and W_2.

2.5.2 The Process Model

$$\frac{dW_1}{dt} = F - B_1 - O_1 \tag{2.54}$$

$$W_1 \frac{dC_1}{dt} = F(C_f - C_1) - O_1 C_1 \tag{2.55}$$

$$W_1 \frac{dh_1}{dt} = F(h_f - h_1) - O_1(H_{v1} - h_1) + Q_1 - L_1 \tag{2.56}$$

$$\frac{dW_2}{dt} = B_1 - B_2 - O_2 \tag{2.57}$$

$$W_2 \frac{dC_2}{dt} = B_1(C_1 - C_2) - O_2 C_2 \tag{2.58}$$

$$O_2(H_{v2} - h_2 + \frac{\partial h_2}{\partial C_2} C_2) = B_1(h_1 - h_2) + Q_2 - L_2 + \frac{\partial h_2}{\partial C_2} B_1(C_2 - C_1) \tag{2.59}$$

Additional relations:

$$h_i = f(T_i) \tag{2.60}$$

$$H_i = f(T_i) \tag{2.61}$$

$$Q_1 = U_1 A_1 (T_{s1} - T_1) = W_s \lambda \tag{2.62}$$

$$Q_2 = U_2 A_2 (T_1 - T_2) \tag{2.63}$$

$$L_1 = h_{w1} A_1 (T_w - T_1) \tag{2.64}$$

$$L_2 = h_{w2} A_2 (T_w - T_2) \tag{2.65}$$

$$O_1 = k_v \sqrt{(P_1 - P_2)} \tag{2.66}$$

$$P_1 = f(T_1) \tag{2.67}$$

$$P_2 = f(T_2) \tag{2.68}$$

Level Controllers:

$$B_1 = B_{1i} + k_{c1}(r_1 - W_1) \tag{2.69}$$

$$B_2 = B_{2i} + k_{c2}(r_2 - W_2) \tag{2.70}$$

Notes:

The current model is different than the original one in the following aspects:

- The overhead flow out of the first effect is defined in the original model by the heat transfer equation, which can be calculates as the amount vaporized at the given temperature of the first effect. In this model it is defined by the pressure difference between the two effects, i.e. (2.66).

- The overhead flow out of the second effect is defined in the original model by the heat balance equation around the second effect (2.59). In this model it is defined as manipulated variable.
- The temperature of the second effect is fixed in the original model. Here it is given by (2.59) after modification as follows:

$$W_2 \frac{dh_2}{dt} = -O_2(H_{v2} - h_2 + \frac{\partial h_2}{\partial C_2} C_2) + B_1(h_1 - h_2) + Q_2 - L_2 + \frac{\partial h_2}{\partial C_2} B_1(C_2 - C_1)$$

- The specific enthalpy of Equation 2.60 & 2.61 are determined from an empirical correlations developed by the author of the original model. Here, common enthalpy correlations for water are used.
- In the current model, water vapor pressure correlations are included in order to calculate the vapor flow rate in (2.66).
- The proportional controllers, (2.69 & 2.70) are used here where indicated to control the liquid level in the corresponding effects.
- The sensitivity of the enthalpy of the solution $\frac{\partial h}{\partial c}$ is given by empirical relation in the original model. Here, it is estimated from enthalpy correlation for brine.

2.5.3 Process Parameters and Variables

Variable	*Description*	*Value*	*Units*
A_1	Heat transfer area of 1st effect		ft^2
A_2	Heat transfer area of 2ed effect	4.6	ft^2
B_1	Bottom stream of 1st effect	3.315	lb/min
B_2	Bottom stream of 2ed effect	1.715	lb/min
C_1	Solute concentration in 1st effect	4.8262	Wt%
C_2	Solute concentration in 2ed effect	9.3307	Wt%
C_f	Solute concentration in the feed	3.2	Wt%
F	Feed flow rate	5	lb/min
h_f	Enthalpy of the feed		Btu/lb
h_i	Liquid enthalpy in the i^{th} effect		Btu/lb
H_i	Vapor enthalpy in the i^{th} effect		Btu/lb

h_{wi}	Heat transfer coefficient in the i^{th} effect	0	Btu/minF ft^2
k_v	Valve coefficient	1.7824	
L_1	Heat loss in the 1^{st} effect	0	Btu/lb
L_2	Heat loss in the 2^{ed} effect	0	Btu/lb
O_1	Overhead stream of the 1^{st} effect	1.7	Lb/min
O_2	Overhead stream of the 2^{ed} effect	1.6	Lb/min
P_1	Operating pressure in the 1^{st} effect	25	psi
P_2	Operating pressure in the 2^{ed} effect	7.5	psi
T_1	Temperature in the 1^{st} effect	225	^oF
T_2	Temperature in the 2^{ed} effect	160	^oF
T_f	Feed temperature	190	^oF
T_w	Wall temperature	77	^oF
U	Heat transfer coefficient	5.2345	Btu/minF ft^2
W_1	Liquid holdup in the 1^{st} effect	30	Lb
W_2	Liquid holdup in the 2^{ed} effect	35	Lb
λ	Latent heat of vaporization of water	948	Btu/lb

2.6 MULTI-STAGE FLASH DESALINATION

2.6.1 Process Description and Flow Sheet

In a typical MSF plant shown in Fig. 2.5 we can distinguish between three basic sections: heat rejection section, heat recovery section and the brine heater. On leaving the first (warmest) rejection stage the feed stream is split into two parts, reject sea water which passes back to the sea and a make up stream, which is then recycled back to the flash section of the last stage. A recycle stream, which is drawn from the last stage, passes through a series of heat exchangers, its temperature rises as it proceeds towards the heat input section of the plant. Passing through the brine heater the brine temperature is raised from the feed temperature at the inlet of the brine heater to a maximum value approximately equals to the saturation temperature at the system pressure. The brine then enters the first heat recovery stage through an orifice thus reducing the pressure. As the brine was already at its saturation temperature it will become superheated for a lower pressure and flashes to give off water vapor. The vapor then passes through a wire mesh (demister) to remove any entertainment brine droplets and onto a heat exchanger where the vapor is condensed and drips into a distillate tray. The process is then repeated all the way through the plant as both brine and distillate enter the next stage which is at a lower pressure. The concentrated brine is divided into two parts as it leaves the plant, a blow-down, which is pumped back to the sea and the recycle stream.

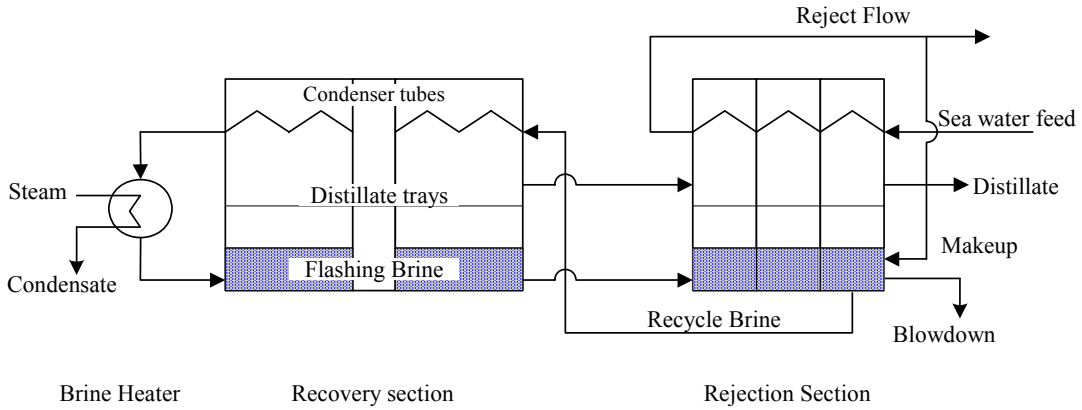


Fig. 2.5 Flow Sheet for the MSF process

2.6.2 The Process Model

The design and analysis of process operation requires the use of a rigorous model of the MSF plant. A first principle model for a 22 stages MSF plant was developed and validated [2]. The specific plant consists of 3 rejection section stages and 19 recover section stages. In the following a summary of the developed model is given. Each plant stage is shown in Fig. 2.5b. The process is therefore defined by nine variables: brine pool height (L_j), brine flow rate (B_j), salt mass fraction (X_j), brine temperature (T_{Bj}), distillate flow rate (D_j), distillate temperature (T_{Dj}), coolant temperature (T_{Cj}), vaporization rate (V_j) and stage pressure (P_j). Furthermore and in order to minimize the size of the model the liquid levels, except that for the last stage, and the temperature dynamics in the distillate tray are not included in the modeling. The dynamics for the salt concentration is also excluded because they have no direct effect on the other process states except through physical properties of the brine such as density and heat capacity. Our simulations revealed that the physical properties vary between +1 and –1 % over the plant temperature range due to changes in salt concentration. The mass holdup of the cooling brine inside the condenser tubes is assumed constant. The following mass and energy equations are written for each stage [2]:

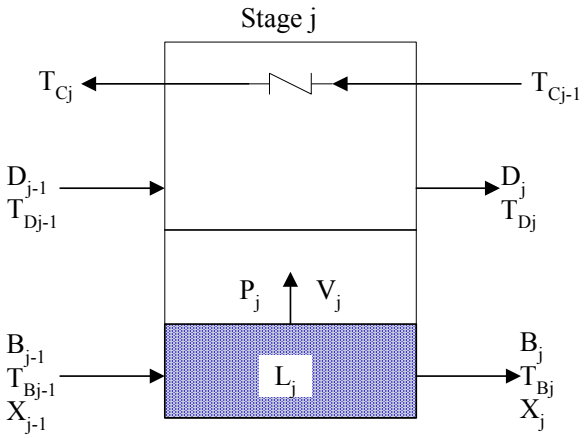


Fig. 2.5b Single Stage

Stage j (except the last stage)

- Mass balance of brine pool

$$\rho_{B,j} A_B \frac{dL_j}{dt} = B_{j-1} - B_j - V_j \tag{2.71}$$

- Energy balance of brine pool

$$\rho_{B,j} A_B L_j Cp_{B,j} \frac{dT_{Bj}}{dt} = B_{j-1} Cp_{B,j} (T_{B,j-1} - T_{B,j}) - V_j (\lambda_{c,j} - Cp_{B,j} (T_{B,j} - T_o)) \tag{2.72}$$

- Mass balance in distillate tray

$$D_j = D_{j-1} + V_j \tag{2.73}$$

- Energy balance of condenser tubes

$$M_{C,j} Cp_{C,j} \frac{dT_{C,j}}{dt} = B_0 Cp_{C,j} (T_{C,j+1} - T_{C,j}) + U_j A_{HC} \Delta T_j \tag{2.74}$$

$$U_j A_{HC} \Delta T_j = V_j \lambda_j \tag{2.75}$$

(for the rejection section the see water feed W_F is used instead of B_0 in (2.74))

Last stage, N

- Mass balance in brine pool

$$\rho_{B,j} A_B \frac{dL_N}{dt} = B_{N-1} - B_N - V_N + W_{mk} - B_0 \tag{2.76}$$

- Energy balance in brine pool

$$\begin{aligned} \rho_{B,j} A_B L_N Cp_{B,j} \frac{dT_{B,N}}{dt} = {} & B_{N-1} Cp_{B,j} (T_{B,N-1} - T_{B,N}) - V_N (\lambda_N - Cp_{B,j} (T_{B,N} - T_o)) \\ & + W_{mk} Cp_{B,j} (T_{C,3} - T_{B,N}) \end{aligned} \tag{2.77}$$

- Mass balance in distillate tray

$$D_N = D_{N-1} + V_N \tag{2.78}$$

- Energy balance of condenser tubes

$$M_{C,N} Cp_{C,N} \frac{dT_{C,N}}{dt} = W_F Cp_{C,N} (T_F - T_{C,N}) + U_N A_{HR} \Delta T_N \tag{2.79}$$

$$U_N A_{HR} \Delta T_N = V_N \lambda_N \tag{2.80}$$

Splitter

$$W_F = Rej + W_{mk} \tag{2.81}$$

Brine Heater

- Energy Balance equation

$$M_{BH} Cp_{B,j} \frac{dT_{B0}}{dt} = B_0 (Cp_{C,1} T_{C,1} - Cp_{B0} T_{B0}) + W_s \lambda_s \tag{2.82}$$

Additional relations:

In the above model equations the brine flow and the brine level in each stage are correlated as follows:

$$B_j = wL_j K_j \sqrt{\rho_{B,j}(P_{j-1} - P_j + \rho_{B,j} g(L_j - Ch_j))} \tag{2.83}$$

Similarly, the distillate flow is correlated to the distillate level as follows:

$$D_j = C_{D,j} \sqrt{\rho_{D,j} g L_{D,j}} \tag{2.84}$$

The temperature difference used in the above energy balances is defined as follows:

$$\Delta T_i = T_{B,i} - 0.5(T_{C,i} + T_{C,i+1}) \tag{2.85}$$

Note that λ_j is computed at T_{Bj} while $\lambda_{c,j}$ is computed at the distillate temperature, which is assumed to be equal T_{Bj} minus the boiling point rise at the j^{th} stage (BPR_j). In the original model [2-4], the physical properties for each stage in the above model, i.e. $\rho, Cp, \lambda, U, M_c, M_{BH}, BPR$ and the vapor pressure (P) are estimated through empirical correlation.. Industrial values were used for the plant design parameters such as C, h, A_B, A_{HC}, A_{HR} and w. Moreover, realistic values for the size and number of tubes were used in computing U, M_c and M_{BH}. The definition of all parameters of the model equations is given in the next table.

2.6.3 Process parameters and variables

Variable	*Description*	*Value*	*Units*
A_B	Cross sectional area for the brine chamber,	1260	m^2
A_{HR}	Heat transfer area for condenser tube at the rejection sections	7919.1	m^2
A_{HC}	Heat transfer area for condenser tube at the recover sections	77314.8	m^2
B	Inter stage Brine flow rate	Calc	ton/min
B_0	Recycle brine flow rates	217.257	ton/min

B_D	blow-down flow rates	29.3	ton/min
C	Orifice contraction coefficient	0.625	
C_D	Discharge coefficient for the distillate tray	1.0	
Cp_C,	Heat capacity for the brine in the condenser tube	Calc.	kJ/kg ^oC
Cp_B	Heat capacity for the brine in the flash chamber,	Calc.	kJ/kg ^oC
D	Distillate flow rate	Calc.	ton/min
g	Gravitational constant	9.8	
h	Orifice height	0.11	m
K	Orifice discharge coefficient	0.68	
L	Brine level	Calc.	m
L_D	distillate level	Calc.	m
M_C	Liquid holdup for the condenser tube	23654.0	Kg
M_{BH}	Liquid holdup for the brine heater	34736.09	Kg
N	Total number of stages	22	
P	*Vapor pressure*	Calc.	Bar
Rej	Reject flow rate,	95.35	ton/min
T_o	Reference temperature	0	^oC
T_F	Sea Water feed temperature	35	^oC
T_B	Brine temperature	Calc.	^oC
T_C	condenser temperature	Calc.	^oC
T_s	Steam temperature	98.1	^oC
T_{B0}	Top brine temperature,	93	^oC
U	Overall heat transfer coefficient for the condenser tube	35	kJ/min ^oC m^2
U_s	Overall heat transfer coefficient for the brine heater	49.5	kJ/min ^oC m^2
V	Vapor rate, ton/min	Calc.	ton/min
X	Salt concentration	Calc/	kg/kg
W_D	distillate product flow rate	19.2	ton/min
W_F	Seawater feed flow rate respectively,	143.816	ton/min
Ws	Steam flow rate	2.452	ton/min
W_{mk}	Make up flow rate,	48.466	ton/min
w	Orifice width	0.5	m

w_N	Orifice width for stage N	3	m
ρ_B	Density of brine	Calc/	kg/m^3
ρ_D	Density of Distillate	1000	kg/m^3
λ	Latent heat for vaporization,	Calc.	kJ/kg
λ_c	Latent heat for vaporization at the distillate temperature	Calc.	kJ/kg
λ_s	Latent heat for steam,	Calc.	kJ/kg

2.7 TWO CSTRs IN SERIES

2.7.1 Process Description and Flow Sheet

Consider two CSTRs in series with an intermediate mixer as shown in Fig. 2.6. A chemical exothermic reaction of the form of: A → B takes place in the liquid phase. The reaction rate is considered to be first order in the reactant species. The feed to the first reactor is a pure species A with volumetric flow rate Q11 at ambient temperature. The liquid inside the reactor converts partially to species B. The heat released from the chemical reaction is removed by cooling water with volumetric flow rate Qcw1 and temperature Tcw1. The outlet of the first reactor has a volumetric flow rate Q1 and temperature T1, and it is mixed with a fresh feed of pure component A. The product of the mixer is then fed to another reactor where the same first order reaction takes place. Water at ambient temperature is also used to cool the second reactor. The outlet flow rates are considered to be driven by the hydrostatic pressure in each tank, thus it related to the square root of the liquid holdup inside the reactor.

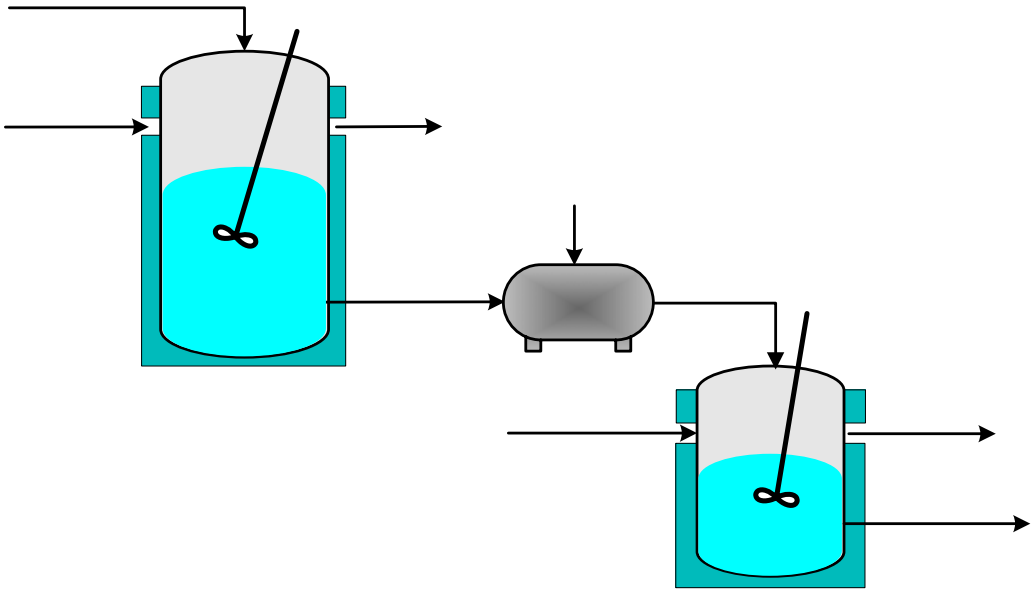


Fig. 2.6 Flow Sheet for two CSTRs in series

2.7.2 The Process Model

The following modeling equations are taken from [7]. State variables are used to describe the process variables. The definition of the states is given in section 2.7.3.

$$Q_{11},\ C_{11},\ T_{11}$$

$$\dot{x}_1 = Q_{11} - K_{v1}\sqrt{x_1} \tag{2.86}$$

$$x_1\dot{x}_2 = Q_{11}(C_{11} - x_2) - K_1 x_1 x_2 \tag{2.87}$$

$$x_1\dot{x}_3 = K_1 x_1 x_2 \Delta H + Q_{11}(T_{11} - x_3) - UA_1(x_3 - x_4) \tag{2.88}$$

$$V_{j1}\dot{x}_4 = Q_{cw1}(T_{cw1} - x_4) + UA_1(x_3 - x_4) \tag{2.89}$$

$$\dot{x}_5 = Q_{12} + K_{v1}\sqrt{x_1} - K_{v2}\sqrt{x_5} \tag{2.90}$$

$$x_5\dot{x}_6 = Q_{12}(C_{12} - x_6) - K_2 x_5 x_6 + K_{v1}\sqrt{x_1}(x_2 - x_6) \tag{2.91}$$

$$x_1\dot{x}_7 = K_2 x_5 x_6 \Delta H + Q_{12}(T_{12} - x_7) + K_{v1}\sqrt{x_1}(x_3 - x_7) - UA_2(x_7 - x_8) \tag{2.92}$$

$$V_{j2}\dot{x}_8 = Q_{cw2}(T_{cw2} - x_8) + UA_2(x_7 - x_8) \tag{2.93}$$

Additional relations:

$$K_1 = k_0 \exp(-E/RT_1) \tag{2.94}$$

$$K_2 = k_0 \exp(-E/RT_2) \tag{2.95}$$

2.7.3 Process parameters and variables

Variable	*Description*	*Value*	*Units*
C_{11}	Feed concentration for component A	20	mole/m^3
C_{12}	Feed concentration for component A	20	mole/m^3
E/R	Activation energy	600	K
Kv_1	Orifice constant for valve 1	0.16	m$^{3./2}$/s
Kv_2	Orifice constant for valve 2	0.256	m$^{3./2}$/s
k_0	Pre-exponential factor	2.7x10^8	1/s
k_1	Reaction rate in CSTR 1		1/s
k_2	Reaction rate in CSTR 2		1/s
Q_{11}	Feed rate for CSTR 1	0.339	m^3/s

Q_{12}	Feed rate for mixer	0.261	m^3/s
Q_{cw1}	Cooling water feed rate to CSTR 1	0.45	m^3/s
Q_{cw1}	Cooling water feed rate to CSTR 1	0.272	m^3/s
T_{11}	Feed temperature to CSTR 1	300.0	K
T_{12}	Feed temperature to CSTR 2	300.0	K
T_{cw1}	Cooling water temperature to CSTR 1	300.0	K
T_{cw2}	Cooling water temperature to CSTR 2	300.0	K
UA_1	Heat transfer coefficient times the heat transfer area for CSTR 1	0.35	m^3/s
UA_2	Heat transfer coefficient times the heat transfer area for CSTR 2	0.35	m^3/s
V_{j1}	Cooling jacket volume for CSTR 1	1.0	m^3
V_{j1}	Cooling jacket volume for CSTR 2	1.0	m^3
x_1	Liquid holdup in CSTR 1	4.4891016	m
x_2	concentration of component *A* in CSTR1	0.0840220	$mole/m^3$
x_3	Temperature in CSTR1	362.99526	K
x_4	Temperature of coolant in CSTR1	327.56042	K
x_5	Liquid holdup in CSTR 2	5.4931641	m
x_6	concentration of component *A* in CSTR2	0.0503550	$mole/m^3$
x_7	Temperature in CSTR2	362.99495	K
x_8	Temperature of coolant in CSTR2	335.44732	K
ΔH	Heat of reaction	5	m^3 K/mole

2.8 FLUID CATALYTIC CRACKING UNIT

2.8.1 Process Description and Flow Sheet

Fluid Catalytic Cracking (FCC) is one of the most important units in the petroleum refining industry for the conversion of heavy gas oil to gasoline and light hydrocarbons. The performance of the FCC units plays major role on the overall economics of petroleum refineries. FCCU receives a number of hydrocarbon feed-streams with different molecular weight distribution from several refinery units and cracks the heavier components in these streams to more valuable lighter ones.

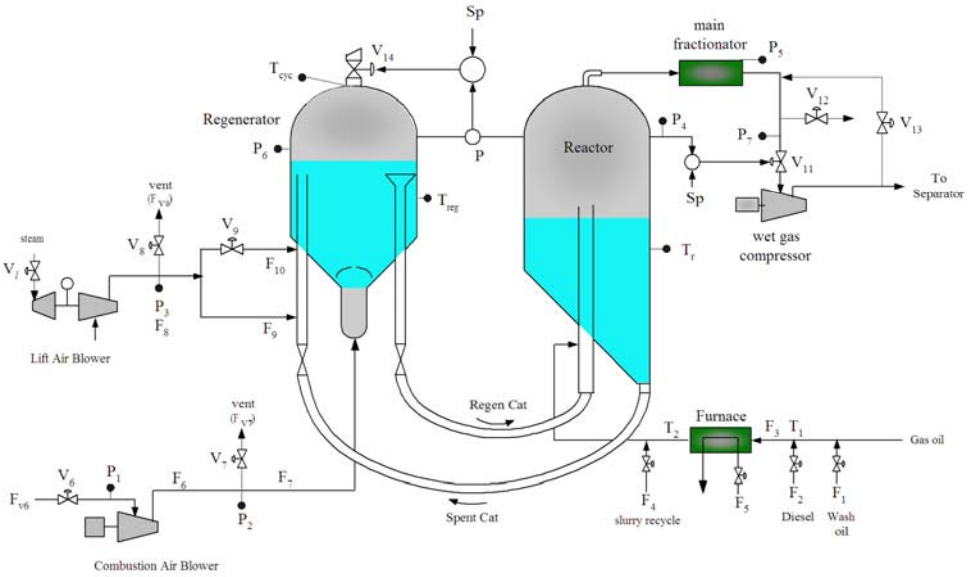


Fig. 2.7 Flow sheet for FFCU process

The FCCU as Shown in Fig. 2.7 consists of two major processes; a reactor and a regenerator. In the reactor, cracking reaction occurs and coke is deposited on the catalyst. The reaction product is passed to the main fractionator for heat recovery and separation into various product streams. The spent catalyst is sent to the regenerator section where the coke is burnt by excess of oxygen to form carbon monoxide and carbon dioxide. The gaseous products of the combustion reaction leave the regenerator section through the cyclones as stack gas. The regenerated catalyst is recycled to the reactor.

The catalyst is transferred back and forth between the reactor and regenerator through the catalyst U-bend. Air is injected at the bottom of the regenerator lift pipe to assist the circulation of the catalyst. Further detailed description of the process can be found in [13].

2.8.2 The Process Model

The process dynamic equation can be found in the previous reference [13]. The model equations are lengthy that we believe not to include them.

2.8.3 Process Parameters and Variables*

Variable	*Description*	*Value*	*Units*	*Type*
C_{co}	Concentration of carbon monoxide in stack gas	37	ppm	State/output
C_{O2}	Concentration of Oxygen in stack gas	0	ppm	State/output
C_{rgc}	Weight fraction of coke on regenerated catalyst	0.0008475	ICFM	State/output
C_{sc}	Weight fraction of coke on spent catalyst	0.0078	ICFM	State/output
Fsucca	Combustion air blower inlet suction flow	49465	ICFM	algebraic
Fsucla	Lift air blower inlet suction flow	11750	ICFM	algebraic
Fsucwg	Wet gas compressor suction flow	19011	ICFM	algebraic
Fsurca	Combustion air blower inlet surge flow	45100	ICFM	constant
Fsurla	Lift air blower inlet surge flow	9571.6	ICFM	algebraic
Fsurwg	Wet gas compressor surge flow	11700	ICFM	constant
F_{air}	Air flow rate into regenrator	75.5375	lb/s	algebraic
F_1	Flow of wash oil to reactor riser	13.8	lb/s	input
F_2	Flow of diesel to reactor riser	0.0	lb/s	input
F_3	Fresh feed flow rate to reactor riser	126	lb/s	input
F_4	Slurry flow rate to reactor riser	5.25	lb/s	input
F_5	Fuel flow rate to furnace	34	lb/s	input

F_6	Combustion air blower throughput,	60.9656	lb/s	algebraic
F_7	Combustion air flow to regenerator	60.9656	lb/s	Algebraic
F_8	lift air blower throughput	14.5719	lb/s	algebraic
F_9	Combustion air flow to regenerator	14.5719	lb/s	algebraic
F_{10}	spill air flow to regenerator	0.0	lb/s	algebraic
F_{11}	Wet gas flow to the vapor recovery unit	1.1275	mole/s	Algebraic
F_{v6}	Flow through combustion air blower suction valve,	60.1634	lb/s	Algebraic
F_{v7}	Flow through combustion air blower vent valve,	0	lb/s	Algebraic
F_{v8}	Flow through lift air blower vent valve,	0	lb/s	Algebraic
F_{v11}	Flow through wet gas compressor suction valve	1.1275	lb/s	algebraic
F_{v12}	Flow through wet gas compressor vent valve	0	lb/s	Algebraic
F_{v13}	Flow through wet gas compressor anti-surge valve	0	lb/s	Algebraic
Lsp	Level of catalyst in standpipe	10.1596	ft	Algebraic
n	Amount of gas	246.25	mole	state
P_{blp}	Pressure at bottom of lift pipe	32.1	Psia	Algebraic
P_1	Combustion air blower suction pressure	14.6421	Psia	State
P_2	Combustion air blower surge pressure	35.2159	Psia	State
P_3	Lift air blower discharge pressure	40.5944	Psia	state
P_4	Reactor Pressure	32.8244	Psia	algebraic
P_5	Reactor fractionation pressure	23.3244	Psia	State
P_6	Regenerator Pressure	29.6707	Psia	State
P_7	Wet gas compressor suction pressure	22.4988	psia	State
T_{cyc}	Regenerator cyclone temperature	1279.2	^oF	State
T_r	Reactor temperature	992.6582	^oF	State

T_{reg}	Regenerator bed temperature	1286.4	^oF	State
T_1	Fresh feed temperature to furnace	460.9	^oF	Input
T_2	Fresh feed temperature to reactor	667.2611	^oF	state
T_3	Furnace firebox temperature	1607.6	^oF	State
V_6	Combustion air blower suction valve position	1		Input
V_7	Combustion air blower vent valve position	0		Input
V_8	Lift air blower vent valve position	0		Input
V_9	Spill air valve position	0		Input
V_l	List air steam valve	0.4313		Input
V_{11}	Wet gas compressor suction valve position	0.95		Input
V_{12}	Wet gas flare valve position	0		Input
V_{13}	Wet gas compressor vent valve position	0		Input
V_{14}	Stack gas valve position	0.612		Input
W_r	Inventory of catalyst in reactor	101696.89	lb	state
W_{reg}	Inventory of catalyst in regenerator	273763.07	lb	State
W_{sp}	Inventory of catalyst in regenerator spent pipe	3200.27	lb	State
ρ_{lift}	Density of catalyst in lift pipe	3.2	lb/ft^3	State

* definition and numerical values of various process parameters are given in the cited reference.

The system has three major PI control loops which are set as:

Control loop	k_c	τ_I
$P_4 \rightarrow V_{11}$	0.1	0.04
$\Delta P \rightarrow V_{14}$	0.08	0.03
$F_{air} \rightarrow V_l$	0.01	0.0025

The process has several equipment and operational constraints as follows:

Flows:

$0 \leq F_1 \leq 17$

$0 \leq F_2 \leq 16$

$0 \leq F_3 \leq 144$

$0 \leq F_4 \leq 10$

$0 \leq F_5 \leq 20$

Valves:

$0 \leq V_i \leq 1;$ $\quad i = 1,14$

Vessels:

P6 < 39.7 psig

P5 < 39.7 psig

Compressor surge limit (level: 10)

Lift air:	Fsuc – Fsur > 0
Combustion air	Fsuc – Fsurg > 0
Wet gas	Fsuc – Fsur < 0

Total combustion in regenerator (level: 9.5)

C_O,sg > 1.5%

$T_{reg} > 1265\ ^oF$

Or

$T_{cyc} - T_{reg} < 20\ ^oF$

$T_{cyc} < 1310\ ^oF$

$C_{CO,sg} < 350$ ppm

Differential pressure: (level 8)

$$-5.0 \leq \Delta P \leq -2.0 \text{ psi}$$

Standpipe level: (level lower limit = 10, upper limit = 8)

$$0 \leq L_{sp} \leq 20 \text{ ft}$$

Riser Temperature: (level 10)

$$T_r \leq 995\ ^{o}F$$

Furnace Temperature: (level 9)

$$T_3 \leq 1700\ ^{o}F$$

CAHAPTER 3: THEORETICAL BACKGROUND

3.1 INTRODUCTION

As mentioned in Chapter 1 and 2, PCLAB deals with exercises that deal with steady state and dynamic analysis of the process. The analysis is essential part of the controller design. In this chapter, the reader will be introduced to some of the terms that will be encountered in most of the exercises to be performed.

The modeling of industrial processes usually starts with a balance on a conserved quantity mass or energy. Mass and energy balances are very important for the engineer to describe the process mathematically. This balance can be written as:

Total flow rate of (mass/energy) into the system	–	Total flow rate of (mass/energy) out of the system	=	Accumulation rate of (mass/energy) within system

(3.1)

As we can imagine, in writing these balances and all other auxiliary equations, we must make use of almost every area of process engineering, such as thermodynamics, heat transfer, fluid flow, mass transfer, reaction engineering, etc. This makes the modeling of industrial processes most interesting and challenging. After characterizing the system by the model, it is then necessary to analyze the behavior of the system in terms of its variables that affect the controlled parameter, or the effect of disturbances that surface to affect the controlled parameter with time. When these variables are known, steps are then taken to keep them in check through the development of control algorithms.

Applying (3.1) usually results in nonlinear ODE equations which can be linearized and converted into transfer function in the form of:

First Order with Dead Time (FODT): $$y(s) = \frac{ke^{-\theta s}}{\tau s + 1} u(s) \qquad (3.2)$$

Second Order with Dead Time (SODT): $$y(s) = \frac{ke^{-\theta s}}{\tau^2 s^2 + 2\zeta\tau s + 1} u(s) \qquad (3.3)$$

3.2 STEADY STATE ANALYSIS

From (3.1), we conclude that the analysis of a process control system requires an understanding of the overall system behavior and the reflection of this behavior in the properties of the system elements. These properties are highlighted during steady state conditions where, the accumulation term in the above equation goes to zero. The state variables do not change with time and the input is then directly related to the output. This leads to the definition of the steady state gain as:

K=Change in output/Change in input **(3.4)**

This relation is useful to determine the effect of manipulated and load variables on the controlled variable of a process. Thus the relation characterizes the effect that a change in an input variable has on an output variable. Implicit in this characterization is the fact that changes in the controlled variable can be calculated by summing up the changes in all the manipulated and load variables multiplied by a constant gain coefficient. This superposition applies only to linear systems. Steady state models are frequently used to determine set points for controllers whose measured variables are dynamic multiple-input computed variables. They are used to adjust the set points of unit operations performance variable..

3.2.1 Steady State Disturbance Sensitivity Analysis

Steady state disturbance analysis is a tool developed by (Yi and Luyben, [17]) to study the controllability and further the control structure of a given process. It is more suitable for nonlinear processes where the effect of process disturbance on the plant steady state operating condition is investigated.

Steady State disturbance sensitivity analysis can be carried out in two modes:

- Open loop
- Closed loop

In the open loop mode, steady state disturbance analysis illustrates how much a chosen process output changes at steady state for various different values of *a given disturbance* when all the manipulated variable are fixed at their corresponding steady state values. In this case, the severity and nonlinearity of the disturbance impact on a certain process output can be investigated. Moreover, one can compare the effect of the designated disturbance on several process outputs and find out which one is the most influenced.

In the closed loop mode, the steady state disturbance sensitivity analysis (*ssdsa*) demonstrates how much a specific input or a group of inputs should change to keep *a controlled output* or group of controlled outputs at its set point at steady state for various values of *a given disturbance* while the other inputs are fixed. Given certain constrains on the inputs, one can select the most appropriate input to regulate a specific output under the influence of a particular disturbance.

The steady state disturbance analysis (SSDA) is obtained by solving the static version of the process model for different values of *a pre-specified disturbance* with *a set of particular outputs* and all the inputs, except those used as manipulated variable, are fixed. The numerical solution is carried out using *fsolve* m-function of MATLAB. Further discussion and analysis of this tool will be discussed in a later chapter.

3.3 DYNAMIC ANALYSIS

The word dynamic conveys the idea of a time variation which can result from a number of unspecified or unknown influences; the word variable simply relates the capacity to vary from these influences. In process control we are interested in those dynamic variables which require regulation in some industrial application. Typical examples are: temperature, pressure, flow rate, level, force, light intensity, and humidity. Dynamic analysis can be performed in open-loop model and closed-loop mode.

3.2.1 Open-loop Dynamic Analysis

In open-loop dynamic analysis no control algorithm is involved. Therefore the dynamic response of a process variable to forced input variation can be studied.

Dynamic response is a measure of the systems reaction, as a function of time, in correcting transient inputs, or adjusting to a new set point. In this case, the dynamic analysis is useful to understand the dynamic behavior of a process and to characterize the process dynamic parameters such as time constant, damping factor, period of oscillation and dead time.

When a FODT process is subject to a step change in its input variable, it responds as shown in Fig. 3.1. The output reaction can be recorded and plotted to infer the system parameter. The figure shows how estimate the gain, time constant and dead time directly from the transient response. This procedure is known as reaction curve method.

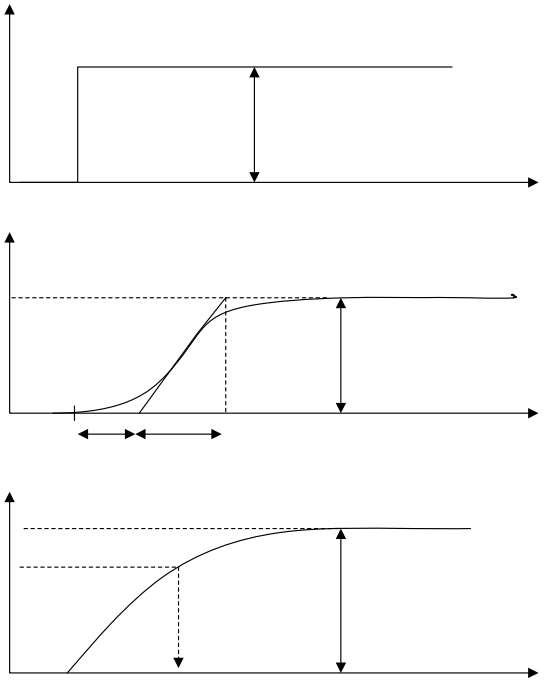


Fig. 3.1: Output response for a step change for first order system, (a) with time delay, (b) without time delay

Accordingly, the time constant is estimated directly from the response curve while the static gain is simply:

$$k = \frac{\Delta x}{\Delta u} \tag{3.5}$$

Similarly, when the input variable is stepped, the output of an under-damped second order system responds as shown in Fig. 3.2. From the recorded data, it is easy to identify the system parameters as follows:

Static gain: $k = \frac{x_{ss}}{\Delta u}$

Damping ratio: solve $\frac{A}{B} = \exp(\frac{-2\pi\varsigma}{\psi})$ for ζ.

Natural period of oscillation: $\tau = \frac{T\psi}{2\pi}$

Where $\psi = \sqrt{1-\varsigma^2}$

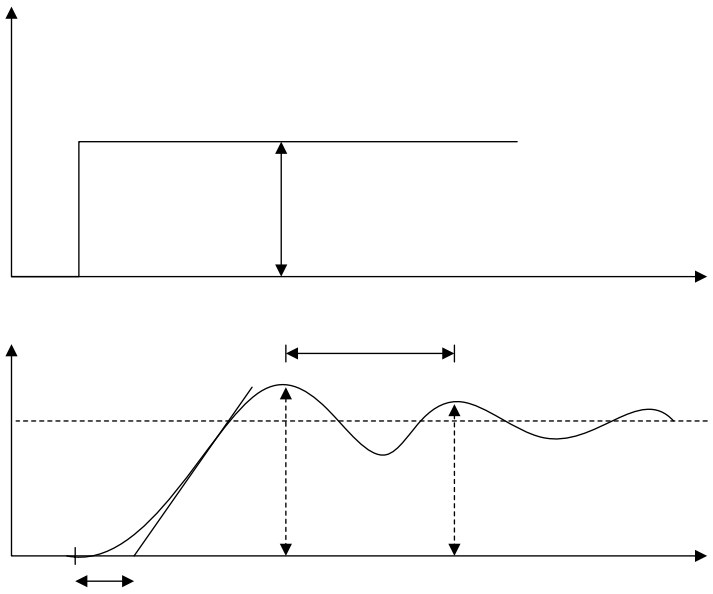


Fig. 3.2: Output response to a step change for second order system

3.2.2 Closed-loop Dynamic Analysis

The evaluation of system performance simply reflects the ability of a process control loop to regulate some dynamic variable in the process. Such regulation is specified through two criteria. The first criterion is the system error of the system in maintaining the controlled dynamic variable at the specific value defined by the set

$$u\,(\mathrm{t})$$

point C_{SP} .The second criterion is the dynamic response of the system to any disturbance of the process or change of the process control loop set point.

The system error is a measure of the inherent error between the system set point value and the actual value of the dynamic variable maintained by the system. The desired value of the dynamic variable in the process is referred to as the set point.

Generally, a set point C_{SP} is expressed with some allowable deviation ± ΔC about the nominal value. Thus, when control has been achieved the actual value of the dynamic variable may be in the range from ($C_{SP} - \Delta C$) to ($C_{SP} + \Delta C$). Of course the wider the allowable deviation the easier the control is to achieve. Note that this deviation can never be less than the inherent system error.

Feedback control loop

The regulation of a specific dynamic variable is usually carried out in feed back fashion as shown in fig. 3.3.

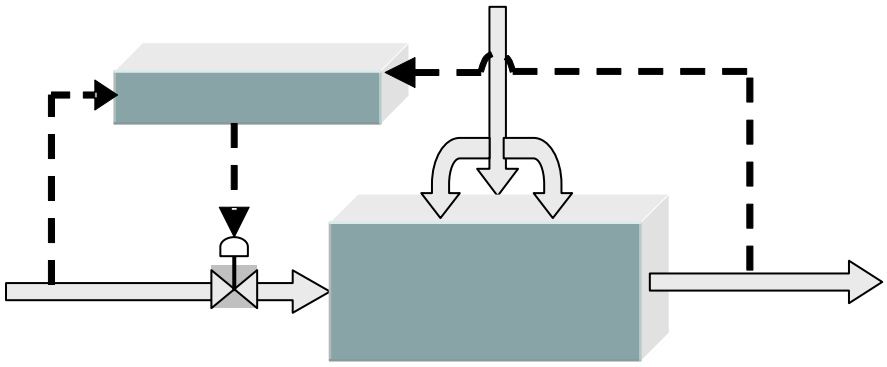


Fig. 3.3 Typical Feedback control configuration

The feedback controller is the component of the feedback loop that issues control commands to the process (via the final control element) based on *e*, the deviation of the process measurement from its desired set point value. The way the controller calculates the type of command signal to be issued for any given value of *e* is what differentiates one controller from another. To consider all dynamic elements of a feedback controller, the following block diagram is usually used:

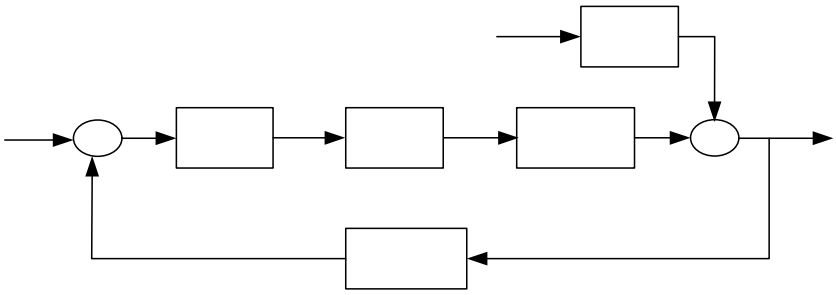


Fig. 3.4 Typical Block diagram for Feedback Control

According to fig. 3.4, the following are defined:

$G_c(s)$ = controller transfer function.

$G_v(s)$ = final control element (valve transfer function).

$G_p(s)$ = process transfer function.

$G_d(s)$ = disturbance transfer function.

$G_m(s)$ = measurement device transfer function.

$C(s)$ = controller signal.

$y_m(s)$ = measured output.

$y_{sp}(s)$ = set point.

$e(s)$ = error .

PID controller

The operating equation for PID controller is:

$$C(t) = K_C\left[e(t) + \frac{1}{\tau_I}\int_0^t e(t)dt + \tau_D\frac{de}{dt}\right] + C_S \qquad (3.6)$$

Taking Laplace transform of the above equation we get:

$$C(t) = K_C\left(1 + \frac{1}{\tau_I s} + \tau_D s\right)e(s) \qquad (3.7)$$

The transfer function for PID controller is:

$$G_C(s) = K_C\left(1 + \frac{1}{\tau_I s} + \tau_D s\right) \tag{3.8}$$

The parameter K_c is called proportional; gain, τ_D is called the derivative time constant and τ_I is the integral time constant.

Feed-forward system

The objective of the feed-forward control is to keep the values of the controlled output as its defined values. It is configured such that predictive action is taken on the controlled variable, thus it does not wait until the effect of a disturbance is already felt in the system but acts appropriately in anticipation of the disturbance. Fig. 3.5 depicts a typical feed forward control element [16].

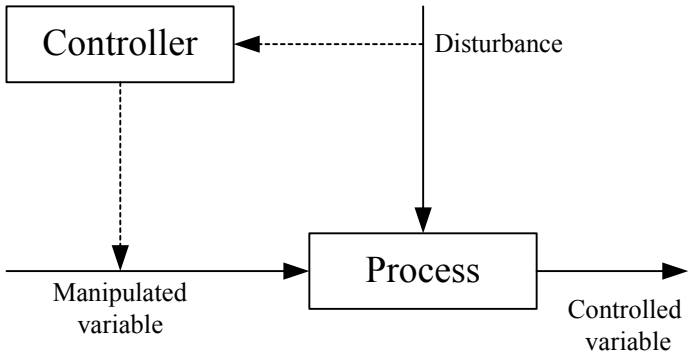


Fig. 3.5 Feed-forward Control configuration

In this control, the direct measurement of a disturbance is used to adjust the values of the manipulated variable to achieve a desired output value. The controller is the active element that receives the information from the measured parameter and takes appropriate action to adjust the values of the manipulated variable.

Consider the open-loop systems to be in the form of:

$$y(s) = G_p(s)u(s) + G_d(s)d(s) \tag{3.9}$$

Let y^{sp} be the desired set point for the controlled variable:

$$y^{sp}(s) = G_p(s)u(s) + G_d(s)d(s) \tag{3.10}$$

The feed forward controller is therefore:

$$u(s) = \left[\frac{1}{G_d}y^{sp} - d(s)\right]\frac{G_d(s)}{G_p(s)} \equiv \left[G_{sp}y^{sp} - d(s)\right]G_c \tag{3.11}$$

And the block diagram with FFC looks:

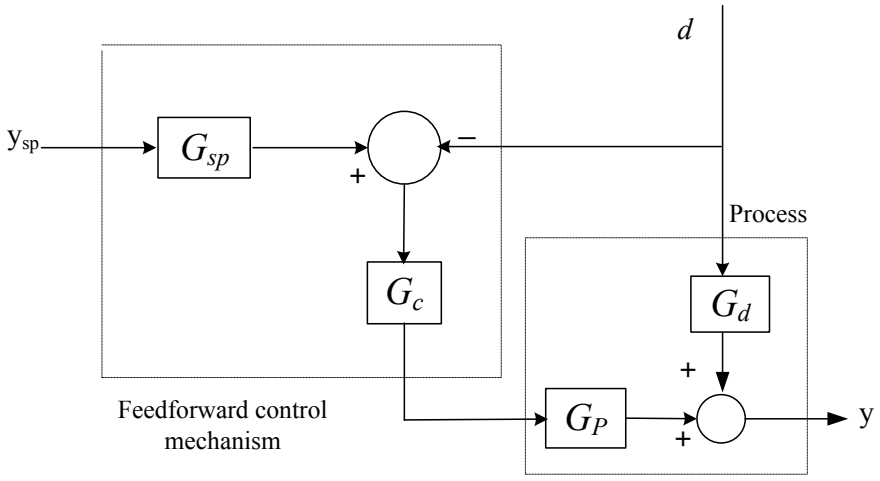


Fig. 3.6: FFC block diagram

Therefore,

$$G_{sp} = \frac{1}{G_d} = \frac{\tau_d s + 1}{k_d} \tag{3.12}$$

$$G_c = \frac{G_d}{G_p} = \frac{k_d}{k_p}\frac{\tau_p s + 1}{\tau_d s + 1} \tag{3.13}$$

Cascade Control

A cascade or multi-loop control is a way of minimizing a disturbance that enters a slow process. It has two sets of controllers, a primary and a secondary. Instead of adjusting the final control element such as a valve, the output of a primary controller is the set point of a secondary control loop. Thus cascade control uses the output of the primary controller to manipulate the set-point of the secondary controller as if it were the final control element. The block diagram for a cascade controller is shown in Fig. 3.7[16].

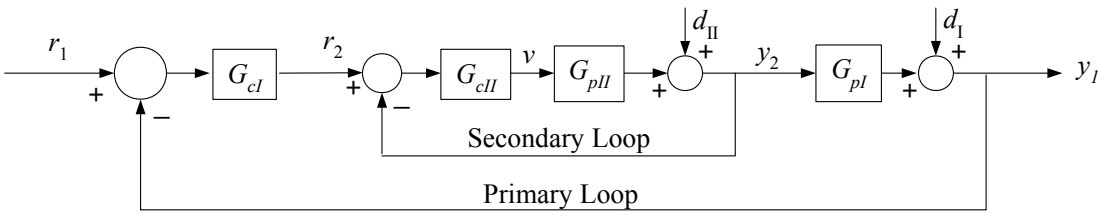


Fig. 3.7 Cascade control configuration

The reasons for using this type of control are as follows:

- Allow faster secondary controller to handle disturbances in the secondary loop.
- Allow secondary controller to handle non-linear valve and other final control element problems.
- Allow operator to directly control secondary loop during certain modes of operation (such as startup).

Requirements:

Secondary loop process dynamics must be at least four times as fast as primary loop process dynamics.

- Secondary loop must have influence over the primary loop.
- Secondary loop must be measured and controllable.

Cascade control can be used when there are several measurement signals and one control variable. It is particularly useful when there are significant dynamics, e.g. long dead times or long time constants between the control variable and the process variable.

CHAPTER 4: STEADY STATE DISTURBANCE SENSETIVITY ANALYSIS

4.1 INTRODUCTION

In this chapter we discuss how steady state disturbance analysis can be implemented on PCLAB. This procedure is very useful to design the appropriate control structure. When controlling a plant or a process with many inputs and outputs, it is usually difficult to optimally pair theses variables into multi single loops structure. SSDA is a tool that can help in this regard, however it can not be implemented on real plant. Instead simulation can be utilized to perform the task.

Referring to the menu of the Evaporator case study in Fig. 1.16, clicking on the "steady state disturbance analysis" button brings up the following window:

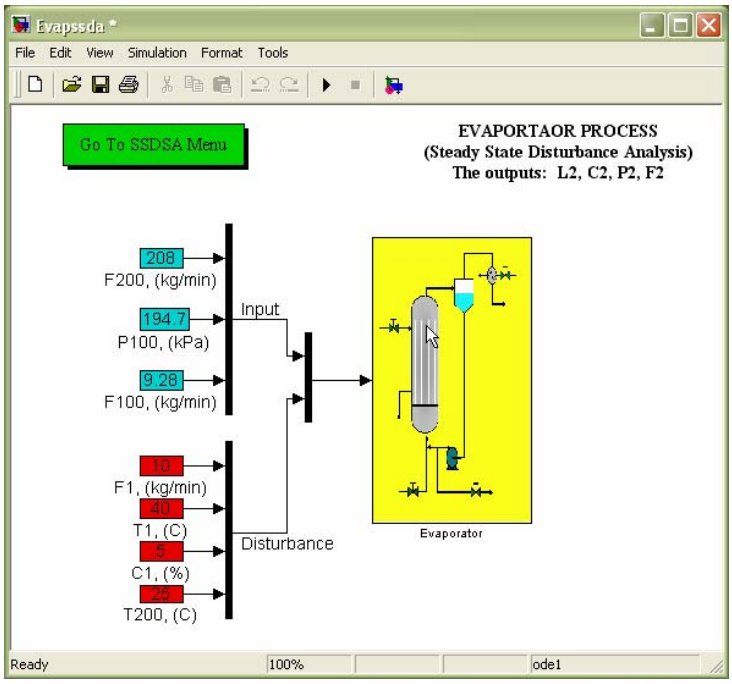


Fig. 4.1 Steady State Disturbance Module for Evaporator Process

Fig. 4.1 shows that the process has three inputs and four possible disturbances. The procedure will focus on investigating the static effect of any disturbance or any combination of disturbances on the process outputs in open-loop mode. This means that the inputs will remain fixed during the test. This is known as the open loop test. It

reveals which process output is affected the most and which one is affected in nonlinear fashion. One can also run the test in closed-loop mode. In this case, an output should be selected as the controlled variable and a corresponding input should be selected to be the manipulated variable. The test will then examine the effectiveness of the chosen input to maintain the controlled variable at its nominal value in steady state when the process is under the influence of a range of disturbance values.

To start the procedure, one can simply click on the green button or the start button on Fig. 4.1. By doing so, the following SSAD menu pops up:

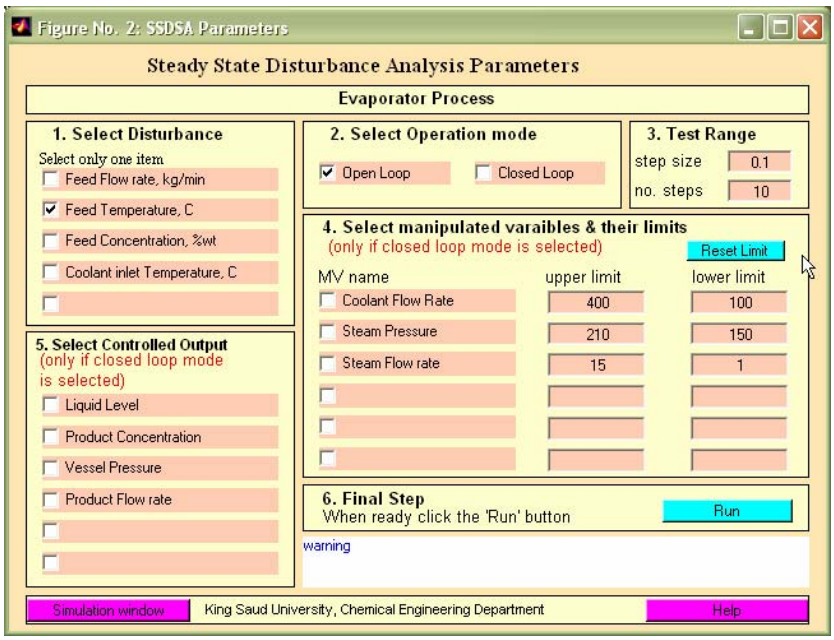


Fig. 4.2 SSDSA menu for the evaporator process.

The menu in Fig. 4.1 has six steps that lead to carry out the SSDA analysis procedure as discussed in the following sections.

4.2 OPEN-LOOP MODE

First the user should select one of the four possible disturbances by marking the appropriate check box. Let us say the feed temperature. Note that the other case studies will have different list of disturbance variables. Next mark the open-loop check box. The third step controls the test range. For example, if the nominal value for the feed temperature is $T_0 = 40\ ^oC$ and using step size of 0.1 and number of steps of 10 then the disturbance value will have the following range during the test:

Increasing the number of steps at the same step size will increase the temperature range to be covered. The above values for the step size and number of steps cover a ±50% range which is good enough from practice point of view. Decreasing the step size will help in producing smoother response curves but it will decrease the overall range. Therefore, if one decreases the step size for better resolution, he should also increase the number of steps to maintain the same operating range. It should be noted though that smaller step size requires higher computational load.

In the open-loop mode, steps 4 and 5 must be by passed. If you by mistake marked one of the boxes in step 4 or 5 , you will receive an error message which will be displayed in the warning box. Now press the run button and look at the results shown in Fig. 4.3.

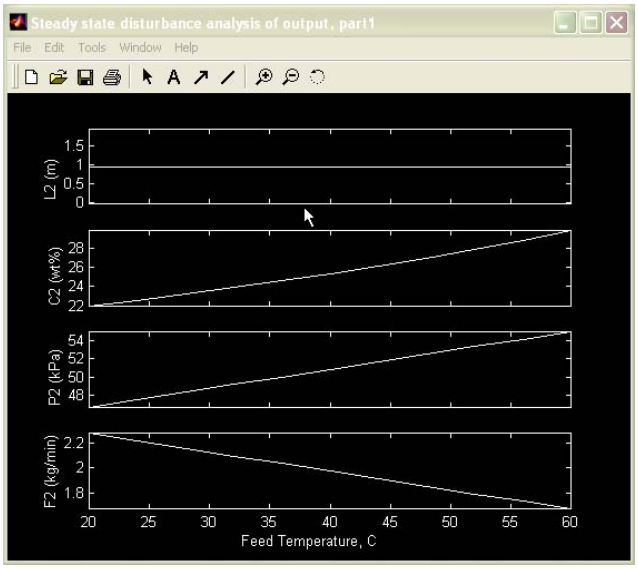


Fig. 4.3 SSDSA results for open loop test.

Fig. 4.3 illustrates how the four main process outputs respond at steady state to changes in the feed temperature from 20 to 60 ^{o}C. It is obvious that the liquid level in the separator unit is not affected by this type of disturbance. Thus in the open loop mode, the user can gain information about the *directional, magnitude* and *nonlinearity* effect of a disturbance. For example as a directional effect, both C_2 and P_2 will increase when the feed temperature increase while the outlet flow rate F_2 will decrease. One can also observe that the solute concentration (C_2) received the highest (*magnitude*) impact. Moreover, all outputs are altered *linearly* with the temperature variation. Thus, the user

can learn how the process operation and product quality may get seriously influenced when the feed temperature is changing freely.

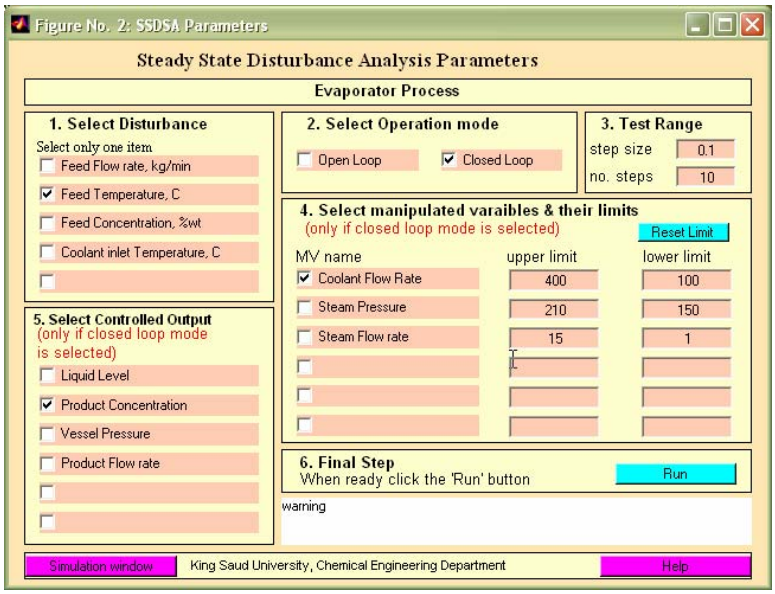
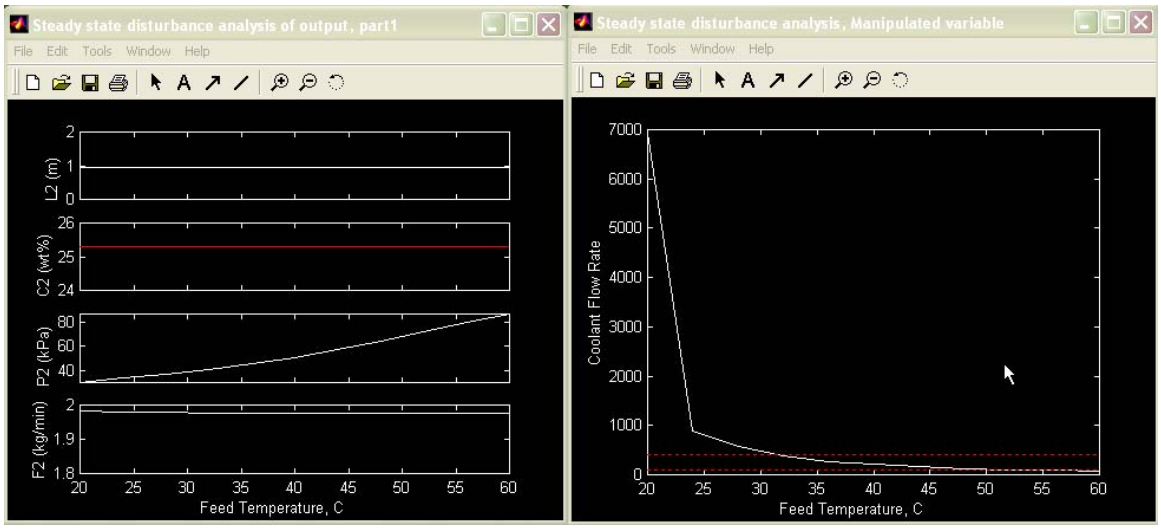


Fig. 4.4 SSDSA menu for the closed-loop test case

4.3 CLOSED LOOP MODE

In this test mode, the user should unmark the open-loop checkbox and mark the close-loop checkbox instead. Furthermore the user needs to specify a controlled variable. Let us choose for example the output concentration C_2. In addition, the user should select one of the manipulated variables listed in Step-Box 4 as the candidate one. Let the candidate manipulated variable be the coolant flow rate. The user has the choice to either change the upper and lower permissible values for the candidate manipulated variable or leave them at their default values. See for example Fig. 4.4. Note that the user can restore the default values for the upper and lower limit any time by simply clicking the “reset limit” button.

After the user finishes marking the required checkboxes in the SSDSA menu, he can simply click the run button in Step-Box 6. The result for the above specification is shown in Fig. 4.5.



(a) (b)

Fig. 4.5 Output responses to disturbance in feed temperature when the coolant flow rate is used as manipulated variable

By inspecting the output response in Fig. 4.5a, we can observe that the controlled variable C_2 is well maintained at the nominal value. Note that a red color is used for the controlled variable to distinguish it from the other uncontrolled outputs. Because the evaporator pressure is not controlled, it increases as expected with disturbance but this time with a larger magnitude. On the other hand, the feedback control helped in making the output flow rate changes slightly with disturbance.

More important is the response of the manipulated variable. The red lines in the right plot of Fig.4.5b shows the upper and lower limits for the coolant flow rate which is set at 400 and 100 respectively. The white line represents the response of the coolant flow rate to disturbances in order to maintain C_2 at nominal value. For large disturbances, i.e. when the feed temperature exceeds 55 ^{o}C, the coolant flow should be reduced slightly below the lower limit in order to reject the effect of the disturbance. However, at feed temperature below 30 ^{o}C, the coolant flow rate must be increased by many folds, especially below 25 ^{o}C, to maintain the required operation. Low feed temperature requires higher steam pressure to provide enough heat of evaporization which in turn increases the process temperature. As a result large amount of coolant flow is needed to absorb the extra heat and to cool down the vapor.

One can conclude that the coolant flow rate is a good manipulated variable for controlling the product concentration at moderate to high disturbances in the feed temperature. However, the coolant flow is recommended at low feed temperature. The user can test other candidate manipulated variable and can also repeat the procedure for the other controlled variables. At the end, the user can build up a satisfactory control structure for the process, i.e. can select the appropriate input-output pairing configuration.

The SSDSA can be studied for each loop individually and consequently the appropriate structure can be determined. This approach however ignores the cross loop interaction. Alternatively, one can conduct the SSDSA method over all process variables and determine the best design structure

CAHAPTER 5: OPEN-LOOP DYNAMIC ANALYSIS

5.1 INTRODUCTION

Of the seven systems that were defined in chapter 2, some follow the response of first order systems, others follow that of second order systems and yet others follow those of higher order systems. In this chapter we will practice the responses of some of these systems to input manipulations or disturbances. The response of an output parameter to changes in an input parameter with time is referred to as system dynamics. Normally the input parameter must behave in a certain defined fashion like step, ramp or sinusoidal.

As discussed in the previous chapter, dynamic analysis can be performed on an open-loop model as well as on a closed-loop model. For systems that mimic first order dynamics, system parameters like gain, time constant and dead time can be computed. These parameters are typical to system response to a step change in the input

Here, it is intended to perform the computations of these parameters using the PCLAB software described in chapter one. This will provide a hands-on application and will enable the reader to calculate the parameters that characterize the system.

5.2 FIRST-ORDER SYSTEM'S DYNAMIC ANALYSIS

In order to run the software, recall how you were instructed to lunch PCLAB in chapter one. Follow the same steps till you arrive at Fig. 1.13. Using the main menu of this figure, choose the Forced Circulation Evaporator to get Fig. 1.16, and then from this menu choose the first item on the menu to get Fig. 5.1.

Fig. 5.1 shows the open-loop configuration of an evaporator process. The blue colors to the left indicate input parameters. There are seven of them as defined in the chapter on modeling. These include temperature, flow rates, pressure and the concentrations defining the process. The red-colored items to the right represent the output parameters that are important to process performance. These four parameters represent the level, pressure, flow and concentration for the evaporator. The yellow portion in the middle, represent the process itself. Observe the green icons on the

top left, one of which can take you back to the menu and the other to the interactive graphic tool. The coloration code is typical for all the case studies available in the software.

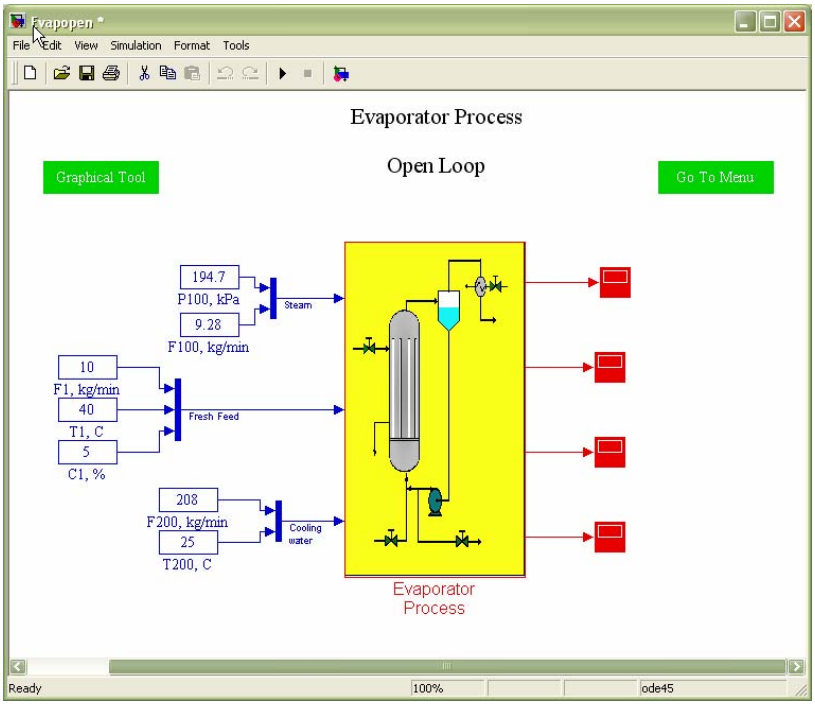


Fig. 5.1 Open Loop Evaporator Process menu

Double clicking on any of the input boxes, will give you access the seven inputs described earlier. For instance, if you need to make a change on the feed flow rate to the evaporator, double click on the third input box, designated by F_1. A Block Parameter window will appear, as shown in Fig. 5.2.

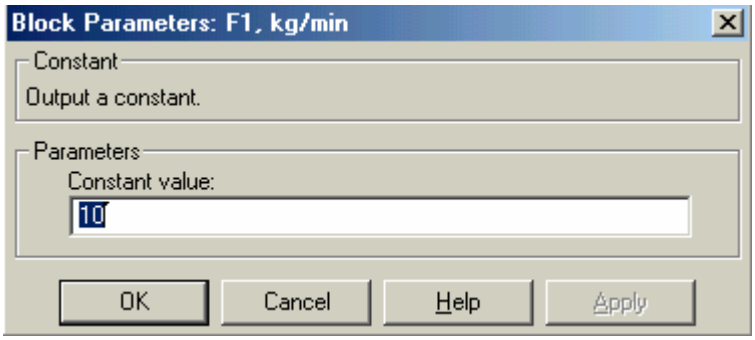


Fig. 5.2 Block Parameter window for feed flow input

Note that the current flow rate is 10 kg/min. To change this number simply use the mouse to highlight the number 10 by right clicking the mouse, hold and drag slightly to the right until the number 10 is fully highlighted, then release the mouse. Now type in the number 11 and press the OK button. This way, a 10% step increase in the inlet flow is attributed to F_1.

After effecting this change, press on the run icon or press simulation and then start. Note that in this case the 300 seconds allotted may not be enough, so click on Simulation then Parameters to get Fig. 4.3. Using the parameter menu option (see Appendix A.3), the stop time of this run should be changed from 300 seconds to 400 seconds. This will enable you to capture enough time for steady state operation. Use the same procedure as in changing the value of the Block Parameter window, change the run time from 300 to 400 seconds.

Now run the simulation by pressing on the start button in the simulation menu. After the 400 seconds has elapsed, the run will be complete. Click on the first output icon for level to view Fig. 5.4. The input has changed step-wise from 10 to 11 kg/min. The output response of the tank to this change is seen to rise steadily to attain a new steady state value. Observe the dead time represented by the flat portion of Fig. 5.4. In this region, the liquid height in the tank does not change with time. In order to read the numerical value of the time during which the system stayed flat, choose the first icon on the top left. Click this icon to allow you enlarge the scale and read the corresponding value on the x-axis as shown in Fig. 5.5. The enlarged portion shows the output response beginning after 10 seconds.

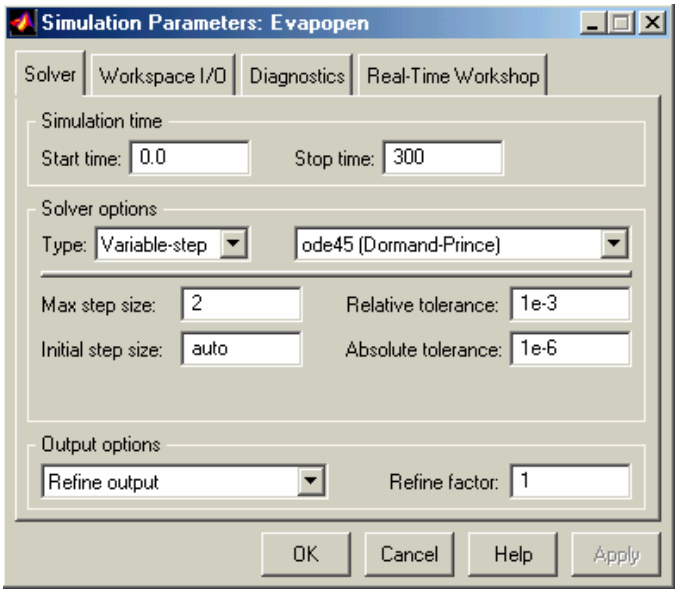


Fig. 5.3 Simulation Parameter window for stop time change

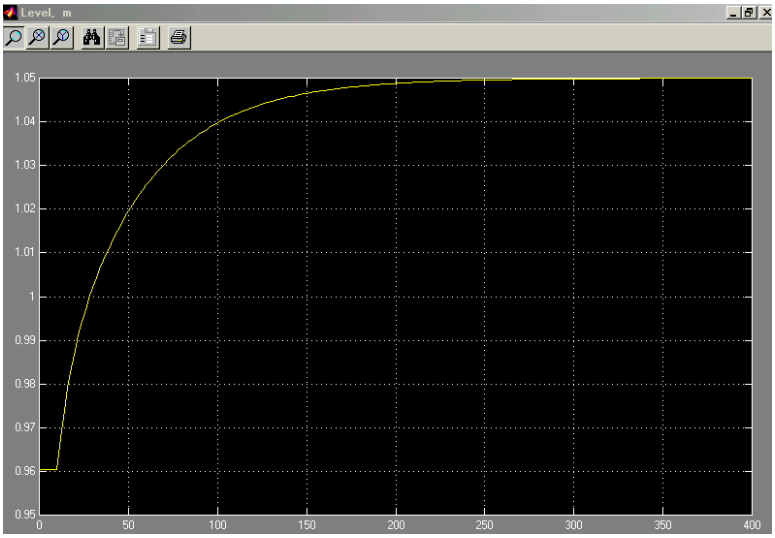


Fig. 5.4 Output showing the height of tank versus time

The region or period is known as the dead time during which the system does not respond to any input change or disturbance. One can then use the plot of Fig. 5.4 to calculate the process gain and the time constant as per (3.2) and (3.3).

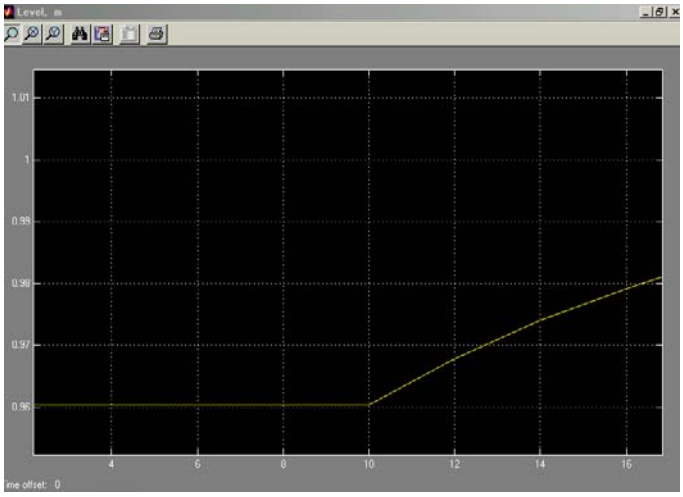


Fig. 5.5 Part of Output showing the height of tank versus time

The gain is defined as the change in output over the change in the input. This ratio is given by (5.1) for the liquid level

$$\frac{L_{ss} - L_0}{F_2 - F_0} \tag{5.1}$$

Where L_0, is the initial level of the tank and L_{ss} the final tank level. F_0 is the initial flow and F_2 the final flow.

This gives a value of 0.089 m-min/kg as the steady state gain. This value of the gain tells us that the for a ten percent change in the input flow rate the liquid level in the storage tank of the evaporator process will change by about 9 percent. Or for a unit step change in the input the tank level will increase by 0.089m.

Similarly, the time constant, τ_p, of the process can also be calculated from the graph. Refer to Fig. 3.1 to refresh your memory on some of these calculations. To do this calculation, the change in the output from the initial steady state to the final steady state value is calculated through (5.2) below.

$$\Delta L = L_{ss} - L_0 \tag{5.2}$$

ΔL, is calculated as (1.05-0.961) to get 0.089m. Now apply (5.3) to get the level of the tank at one time constant.

$$L_{\lambda P} = L_0 + 0.632 * \Delta L = 1.017248 \quad (5.3)$$

Double click on the first red output icon to revisit Fig. 5.4. Then use the icon with binoculars to auto-scale your plot. Now use the three zoom options to the top left of your plot to vary the y-axis until you read the height of 1.017m. Note the corresponding reading on the x-axis as 46.5 seconds. This gives you the value of λ_p for the process.

Now look at the output concentration icon by double clicking on it. Enlarge the graph to obtain Fig. 5.6. Notice how the concentration response falls rapidly to a new level of concentration as against the rise in the level of the tank. Using the same methodology, the gain for the concentration icon was calculated as –7.03 %-min/kg. This means for a unit step change in the input flow, the concentration falls seven percentage points. From this it is clear that process gains can be negative. The time constant was calculated as 17.05 seconds. The concentration response is therefore seen to be 2.7 times faster than the level response.

Table 5.1 Calculated values of the characteristic parameters for positive step change

	K_p	τ_p	t_d
Level	0.089	46.5	10
Concentration	-7.03	17.05	10
Pressure	-0.4472	46.2	10
Flow	0.033	37.5	0

Table 5.1 was generated for all the four output parameters that were calculated using the procedure above. The graphs of the outputs were plotted separately and the

values of K_p, , τ_p and t_d were all calculated from the graphs based on 10% change in the input flow.

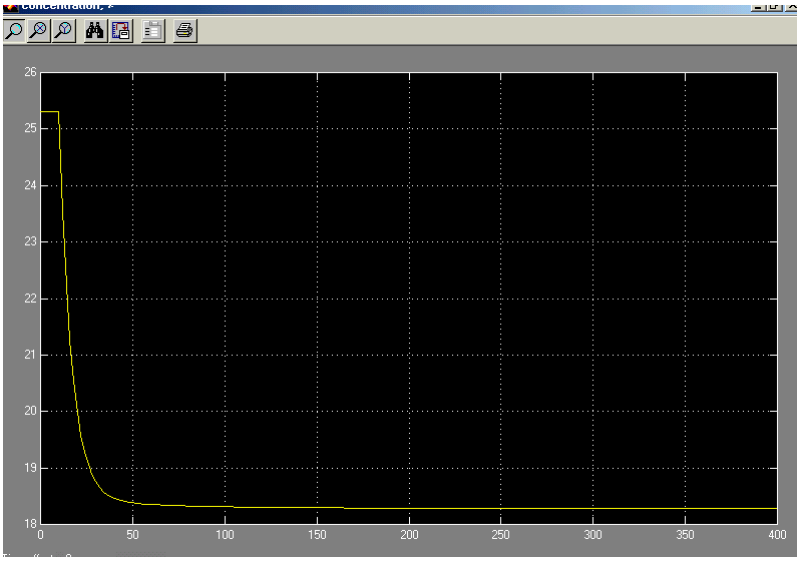


Fig. 5.6 Output showing the change in concentration with time

If K_p is large the system becomes very sensitive and a small change in the input forcing function will result in a very large response. The output concentration has the largest gain and therefore a small change in the input flow will result in a big change in the concentration. On the other hand, The outlet flow has the smallest gain of 0.033, so if large changes are made in the input flow, the outlet flow will not be unduly disturbed. It is possible for the gain parameter to be negative. If it is negative, then the next steady state value will be lower than the initial steady state and the curve will be seen to fall to the new value as in the cases of pressure and concentration. However, if the gain is positive, then the response curve rises and settles at a higher new value.

The time constant, is a measure of how fast a particular parameter responds to an input forcing function. The concentration parameter has the smallest time constant and therefore it is the most sensitive to changes in the inlet flow. The pressure and the flow both respond within the same magnitude of time.

All the parameters show the same value for the dead time except the outlet flow. This is the time it takes before the parameters start or begin to respond to the forcing function. Probably, this dead time is inherent in the system.

The analysis can also be done by storing the raw data to a workspace which could be used in other software like EXCEL or the MATLAB itself. To achieve this, double click the output icon whose data you wish to store before starting the run. For instance, if you wish to store the data of the level in the tank, then double click the first output icon of Fig. 5.1. Doing so will open for you the window similar to Fig. 5.4 but without any plots. Then click on the sixth icon on the top left (Properties). Next, click on the Data History icon to get Fig. 5.7.

Now, mark the box of 'Save data to workspace' and give it a variable name. Choose "Matrix" for Format box. Click on 'Apply' or 'OK' to continue. Make your run or simulation as described earlier and continue. The stored data can be viewed on MATLAB command window by typing the variable name at the prompt option of MATLAB. The data will be stored in MATLAB workspace as a matrix under the chosen variable name. the data can be saved for further analysis or plotting.

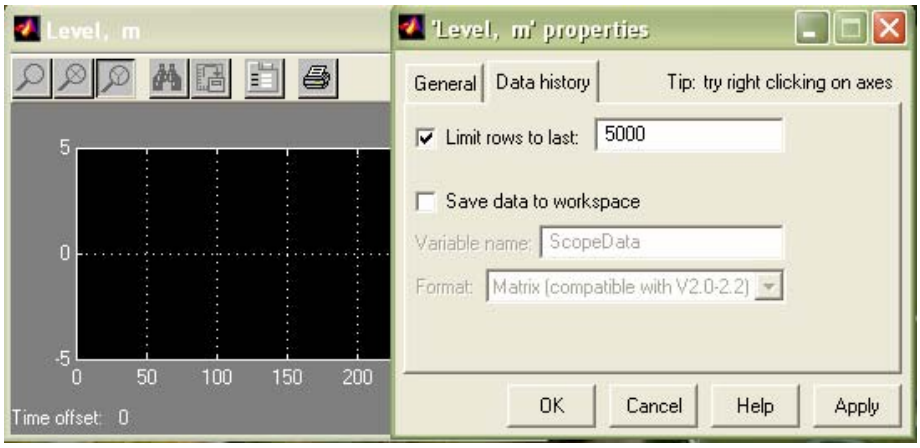


Fig. 5.7 Level Properties window for saving data to a workspace

5.2.1 Nonlinearity Analysis

Approximating the process dynamics by a first order system is based on that assumption that the physical process responds linearly to input variation. However,

this is not true for most chemical processes except at a very narrow operating region. Therefore, it is necessary to use small process perturbation in order to obtain faithful process parameters, i.e. static gain and time constant. However, the input perturbation should be large enough to create reasonable signal to noise ratio. In this sense, the appropriate procedure for identifying the process parameters is a trial-and-error. This will be explained in the next paragraphs.

Now, let us see what happens when the step change is –10%. This is done the same way by invoking the Block Parameter of F_1 and changing the flow rate from 10 to 9 kg/min. Fig. 5.8 is the graph for the response of the tank to the input forcing function. From the graph, there is an immediate inverse response before the curve picks up to respond normally. Similar to the previous paragraph, the parameters of this run are captured and listed in Table 5.2

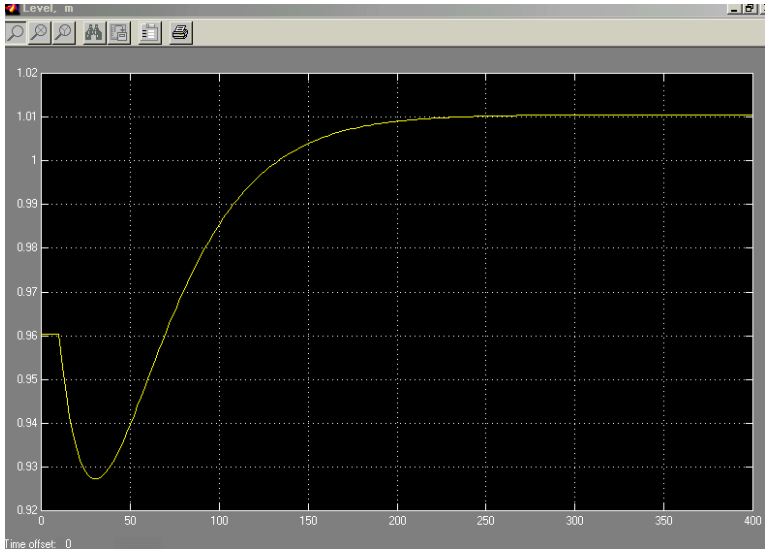


Fig. 5.8 Output showing the height of tank versus time

Table 5.2 Calculated values of the characteristic parameters for negative step change

	K_p	τ_p	t_d
Level	-0.05	111.2	10
Concentration	-20.1	28.8	10
Pressure	0.26	116	10
Flow	-0.0235	118	0

Table 5.2 still shows the concentration output parameter to be the most sensitive with a time constant of 28.8 seconds. The gain is seen to be very high reaching 20.1 % for a step change of –10 %. The gain and the time constant parameters in Table 5.3 are different from those in Table 5.2 and 5.1 for an input step change of 10% and –10%. It should be noted that the dead time was not affected by the step change because it is intrinsic property of the process. Moreover, the dead time incorporated by ad hoc method.

For linear systems, the process parameters must have the same absolute magnitude for the same magnitude of step change. Therefore the systems dynamics for 10% step change do not exhibit linearity in that region. Since linearity is the basis of these calculations, further tests need to be carried out to establish the existence of linearity. One can also check the process linear behavior through testing the plant for step changes of specific proportional magnitudes. In this case, the calculated process parameters should have the same proportional magnitudes. Otherwise the process is said to behave nonlinearly.

Table 5.3 Comparison of the values of the characteristic parameters

	K_p	τ_p	t_d
+10% step change	0.089	46.5	10
-10% Step Change	-0.05	111.2	10

In order to establish this linearity, the step changes were reduced by ten orders of magnitude. Instead of an increment of 1 kg/min on the flow rate, an increment of 0.1 kg/min was made. The module is simulated and the corresponding process parameters were estimated as shown in Table 5.4. The step change was then doubled to an increment of 0.2 kg/min. using the same procedure outlined above, Table 5.4 was produced. In this region an approximation of linearity can be assumed because the values of the gains are close to each other at 0.058 and 0.06 m-min/kg respectively, while the time constants are also close to 30 seconds.

Table 5.4 Calculated values of the characteristic parameters for 1% step change.

	Step Change 0.1 kg/min	Step Change 0.2 kg/min
L_0	0.9604	0.9604
L_{ss}	0.9662	0.9729
ΔL	0.0058	0.0125
$0.632\Delta L$	0.00367	0.0079
$L_0 + 0.632\Delta L$	0.9641	0.9683
L_τ	0.9641	0.9683
τ	27.6	30
K_p	0.058	0.06

5.3 SECOND ORDER SYSTEM DYNAMICS

Let us now look at the open-loop response of second order systems. These systems are known to have differential equations of the form given in (3.3). The Laplace transform of this differential equation yields a second order transfer function given by the equation below.

$$\frac{y(s)}{u(s)} = G(s) = \frac{Kp}{\tau^2 s^2 + 2\zeta\tau_p s + 1} \tag{5.4}$$

Where:

$y(s)$ is the system output

$u(s)$ is the input forcing function

K_p is the system gain

τ_p is the natural period of the system

ζ is the damping coefficient

Second order systems occur in nature. Such systems are said to be inherently second order. They can be derived from a multi-capacity system, such as two first

order systems in series through which material or energy flow. Many closed-loop systems also exhibit second order behavior.

In this chapter, the dynamic behavior of these systems is to be studied based on step changes in the input forcing function. When there is an input forcing function, the response of these systems follows that of the transfer (5.4). The key parameter of this equation is the damping factor. Depending on the value of the damping factor, the system can be overdamped, critically damped or underdamped.

Fig. 3.2 shows a typical output response of an under-damped second order system to a step change in the input. From the data generated by simulating the system, its characteristic parameters can be calculated as shown in the Figure.

In this case, the Polyethylene Reactor example is to be used to make the necessary parametric calculations. To do this, revisit Fig. 1.13 and choose the Polyethylene example on the menu by clicking on it to get Fig. 5.9. The figure shows the Polyethylene Process Menu. Choose the second item on the menu (Polyethylene (stable)) by simply clicking on it. This leads to the process flow sheet of the polyethylene process as shown in Fig. 5.10. Remember, the coloration code discussed earlier still applies.

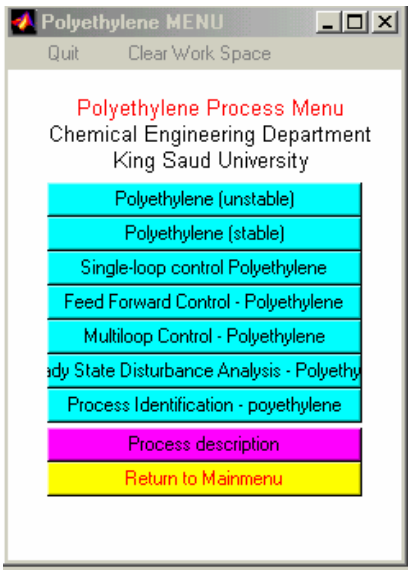


Fig. 5.9 Polyethylene Process Menu

Double click on the input parameter of the recycle flow rate F_g, to obtain the Block Parameter icon. Then increase this recycle flow from 8500 to 9020 moles/s. After this is done, close the Block parameter icon for the recycle and double click that for hydrogen flow, F_H to obtain Fig. 5.11. Once this is displayed on your screen use the keyboard to enter 2.2 moles/s instead of the 1.16. Press the enter button for the system to accept this entry. Now, run the simulation as the same way it was done in the first part of the chapter. When the run button is pressed, PCLAB will run your simulation based on the values of the new parameters that were entered.

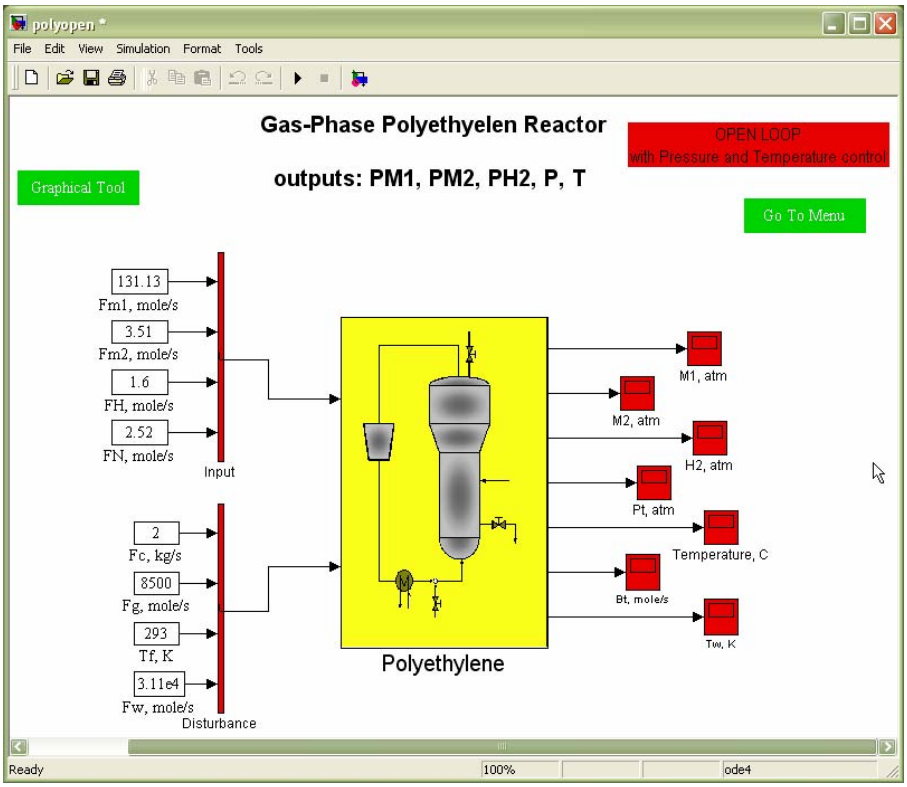


Fig. 5.10 Process Flow Diagram of the Polyethylene system

After the run is complete, double click on the last red icon to the right of your screen, the temperature T, to get Fig. 5.12. The graph shows the plot of output bed temperature versus time for a typical second order system. The nature of the graph resembles that of Fig. 3.2. Applying the definitions used in the Figure, A is found by subtracting the steady state height (T) from the maximum reading of 122.65 ^oC as shown in Table 5.5. B is similarly calculated as the ultimate value of the response. Then the ratio of A/B gave the overshoot. The period is also calculated by taking the time elapsed between two consecutive crests. Using the equations and the appropriate figures of Chapter 3, Table 5.6 was generated. The overshoot is the ratio of the maximum amount by which

the response exceeded its ultimate value to the ultimate value of the response. It is related to the damping factor through the equation shown in the Table. It increases with decreasing ζ and as ζ approaches 1, the overshoot takes the value of zero and the system response is described as being critically damped.

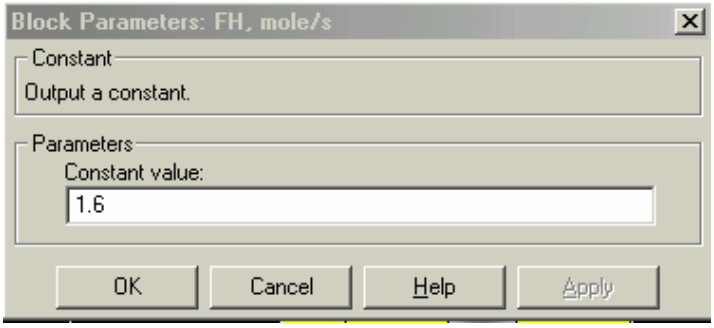


Fig. 5.11 Block Parameter for hydrogen flow, FH moles/s

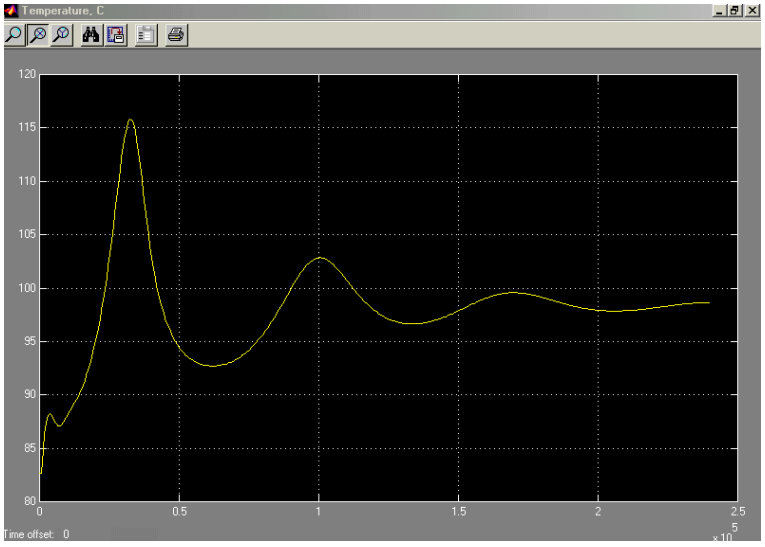


Fig. 5.12 Simulation Results of the Open Loop Polyethylene Reactor.

The system seems to have a very large time constant of 11515 seconds. The static gain of 26.5 seems to be equally large. This shows that the system has very large capacitance. Good design engineers are able to reduce these parameters so that the system can have quick response.

Other parameters that can be calculated from the graph are the rise time and the response time. Since the system took a response time of 2.29×10^5 seconds to reach its ultimate value with equally large value of the rise time (found to be 22,200 seconds), then the system is expected to have large capacitance. The system seems to have a very large time constant of 11515 seconds. The static gain of 26.5 seems to be equally large.

This shows that the system has very large capacitance. Good design engineers are able to reduce these parameters so that the system can have quick response.

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Table 5.5 Calculated characteristic parameters for the Polyethylene Process

Parameter	Equation	Value from Graph
A	122.65-99.9	22.75
B	99.9	99.9
OVERSHOOT	A/B	0.2277
DECAY RATIO	$(\text{OVERSHOOT})^2$	.0519
T_2	$8.76*10^4$	87600
T_1	$1.52*10^4$	15200
T	T_2-T_1	72400
T_n	$\frac{T\psi}{2\pi}$	11515
ζ	$\exp(\frac{-2\pi\varsigma}{\psi})$	0.03621
τ	$\frac{T\psi}{2\pi}$	11515
K_p	$\frac{x_{ss}}{\Delta u}$	26.5

Remarks:

- The green box labeled “Go to menu” shown in the Simulink Module can be used to switch between the active Simulink module with the current active sub menu.
- The green box labeled “graphical tool” shown in the Simulink Module can be used to trigger the plotting forum. The usage of the plotting forum is discussed in Appendix A.

CHAPTER 6: CLOSED-LOOP DYNAMICS

6.1 INTRODUCTION

A closed-loop system normally has a controller together with other hardware components, but the key component here is the controller. The closed-loop response also depends on the transfer function of the system and on the nature of the change. Normally the change is divided into two, namely, set point and disturbance. If there is a set point change, the feed back controller acts in such a way as to keep the ultimate response as close as possible to the changing set point. On the other hand, if there is any form of disturbance but the set point remains the same, then the feed back controller tries to eliminate the impact of the disturbance or load changes and keeps the ultimate response at the desired set point. However, the presence of other hardware components could lead to oscillatory behavior and unstable systems. Therefore it is important to design controller systems that will eliminate all the instability.

6.2 CONTROLLER TUNING

Performance of feedback controllers depends on the values of their chosen parameters. If these parameters are properly chosen, they offer the highest flexibility to achieving the desired controlled response and stability. The process of choosing these parameters is known as controller tuning. There are three general approaches that are used in the controller tuning process. These are; Time integral performance criteria, Use of simple criteria such as one quarter decay ratio and Semi-empirical rules, which have been proven in practice. In this section use is to be made of the semi-empirical rules to tune the controller parameters on some of the examples in the exercises. The two main semi-empirical rules used here are those of Cohen and Coon also known as the process reaction curve method derived from open-loop systems. The other is the Ziegler and Nichols method derived mainly from closed-loop systems.

The Cohen-Coon tuning method requires the response of an open-loop system to an input step change. The response generates a process reaction curve, which can adequately be represented by a first order system with dead time. From the generated curve, the static gain, the time constant and the dead time can all be estimated and used in the rules summarized in Table 6.1 to find the controller parameters.

The Ziegler-Nichols is based on frequency response analysis. It requires the generation of two key process parameters from a closed-loop system with a proportional controller. The two parameters are obtained by increasing the gain of the proportional controller until the closed-loop system exhibits sustained oscillations of constant amplitude. The period of these oscillations is defined as the ultimate period (P_u). The controller gain at which these oscillations occur is referred to as the ultimate gain (K_u). It is these two parameters that are used in conjunction with the equations of Table 6.2 to arrive at the tuned parameters for use in the controller.

These two semi-empirical methods will be used to calculate Controller parameters for a PID controller on the Evaporation process. The values obtained will be compared with each other and used to run the processes mentioned in order to draw conclusions on the superiority (if any) of one method over the other.

Table 6.1 Cohen-Coon Formulas.[16]

Controller Type	Controller Gain Kc	Reset time τ_I	Derivative Time τ_D
P	$K_c = \left(\frac{1}{K_p}\right)\left(\frac{\tau_p}{t_d}\right)\left(1+\frac{td}{3\tau_p}\right)$		
PI	$K_c = \left(\frac{1}{K_p}\right)\left(\frac{\tau_p}{t_d}\right)\left(\frac{9}{10}+\frac{td}{12\tau_p}\right)$	$\tau_I = t_d\left(\frac{30+3t_d/\tau_p}{9+20t_d/\tau_p}\right)$	
PID	$K_c = \left(\frac{1}{K_p}\right)\left(\frac{\tau_p}{t_d}\right)\left(\frac{4}{3}+\frac{td}{4\tau_p}\right)$	$\tau_I = t_d\left(\frac{32+6t_d/\tau_p}{13+8t_d/\tau_p}\right)$	$\tau_D = t_d\frac{4}{11+2t_d/\tau_p}$

Let us now revisit Fig. 1.13. Using the same steps described in chapter 5, the open-loop response for the evaporator process could be used to obtain Table 5.5 and the values of this table (for the gain, time constant and the dead time) are plugged into the equations of Table 6.1 to obtain the Cohen-Coon controller parameters. Another way of obtaining the open-loop parameters of Table 5.5 is to run the single closed-loop module in an open-loop mode fashion as discussed in the following sections.

Table 6.2 Ziegler-Nichols Formula [16]

Controller Type	Controller Gain Kc	Reset time τ_I	Derivative Time τ_D
P	$K_c = 0.5K_u$		
PI	$K_c = 0.45K_u$	$\tau_I = \frac{P_u}{1.2}$	
PID	$K_c = 0.6K_u$	$\tau_I = \frac{P_u}{2}$	$\tau_D = \frac{P_u}{8}$

Revisit Fig. 1.13 of chapter one. Using the main menu of this figure, choose the Forced Circulation Evaporator to get Fig. 1.16, and then from this menu choose the second item (Single-loop control Evaporator) to get Fig. 6.1.

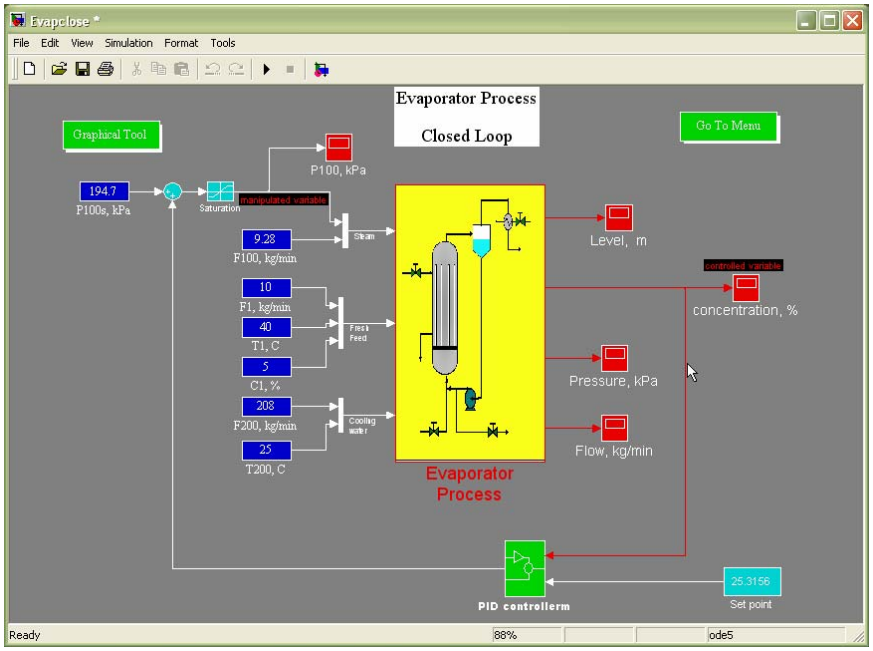


Fig. 6.1 Process flow sheet of closed-loop evaporator process.

This figure represents the closed-loop set up for the evaporator process. Notice the addition of two new color codes different from the previous figures. The light green color code is the designation for the controller, in this case using a PID controller to control the outlet concentration of the evaporator process. The manipulated variable in this case is the feed steam pressure. The light blue color code represents that for the set point. The loop pairing is chosen for demonstration purposes. The user can simply reconfigure the control loop to link any output to any input presented in the module.

6.2.1 Selecting the Coon and Cohen Parameters (CC)

Should there be a need for the open-loop parameters, the reader can simply run the simulation open-loop mode. To do so, double click the PID controller Box (green Box) and enter zero for all controller parameters as shown in Fig. 6.2.

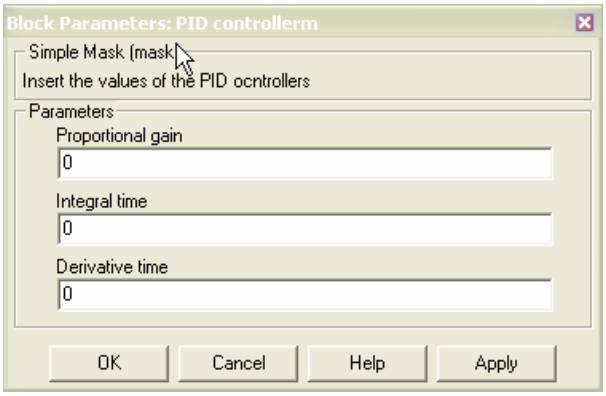


Fig. 6.2 Block Parameter for changing the PID controller settings

Afterward, inset a step change in the desired manipulated variable, say the steam pressure. The step change should be well known value as discussed in chapter 5. Here, the steam pressure was stepped by 1 kPa. Now, run the simulation the usual way and observe the simulation results for the controlled variable which is the product concentration. The plot of the response of the controlled variable is shown in the designated scope box given in Fig. 6.3. From this response calculate the open-loop parameters as discussed in chapter 5. The parameters calculated for this run, i.e. k_p, τ_p and t_d, are given in Table 6.3. Using these values in the Coon and Cohen formula of Table 6.2, the PID controller settings are calculated in listed in table 6.3

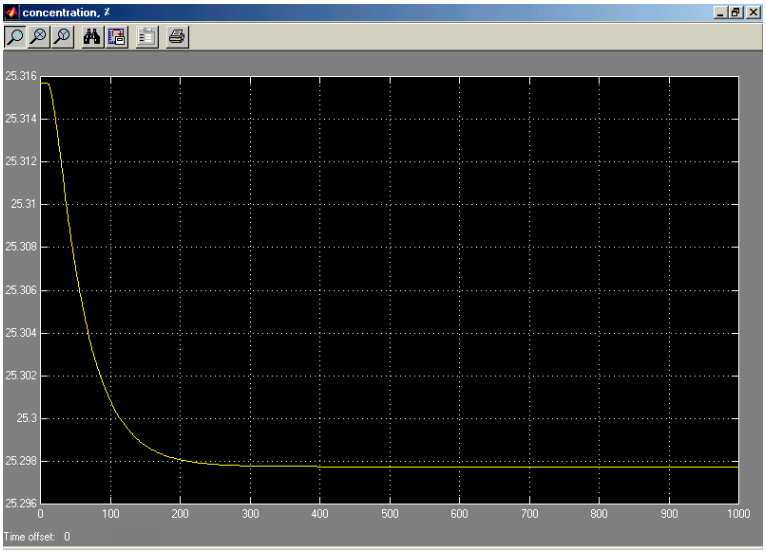


Fig. 6.3 Plot of Concentration with time for a unit step change in P_{100}

6.2.2 Ziegler and Nichols Method (ZN)

To obtain the parameters for the Ziegler-Nichols controller settings it is necessary to make two changes on the closed-loop process. First double click on the set-point icon of Fig. 6.1 to obtain Fig. 6.4. Change the set point value by highlighting the current value and type 26.3156 (set point change of unity).

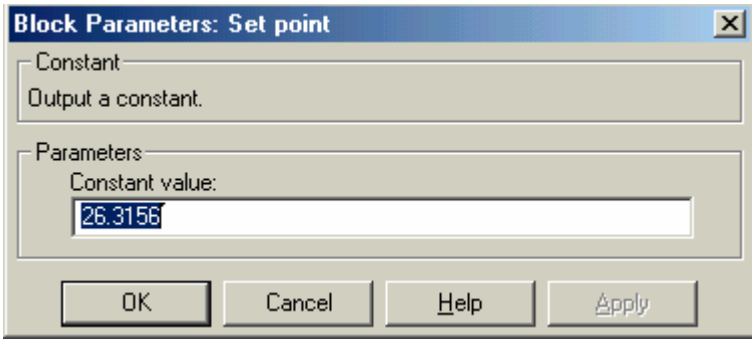


Fig. 6.4 Block Parameter for changing the set point

This way a set point change is introduced. Press the enter button to apply this change to the process. Now double Click the block parameter for the PID controller icon. For the proportional band enter the value –320 by highlighting the current value and typing on it. Then change the settings for the reset time and the derivative time to read zero as shown in Fig. 6.5.

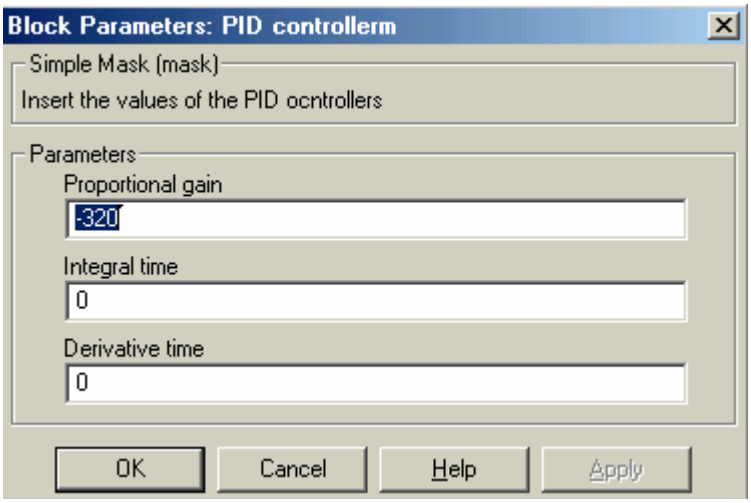


Fig. 6.5 Block Parameter for PID controller

Again, run the simulation the usual way and observe the result shown in Fig. 6.6. It portrays the response of a set-point change in concentration. As the concentration rises to the new set point, the response starts to oscillate, but slowly the oscillation dies away and begins to stabilize like a typical second order system discussed in the previous chapter. Since the oscillation is not a sustained one, it is necessary to make a second guess of the proportional gain. To do this, again double click the light green PID Controller representation to get Fig. 6.5. When you obtain the Block Parameter icon, change the value of the proportional gain from –320 to –350. Press the OK button and rerun the simulation. Observe the response on your screen similar to Fig. 6.7. Notice that this time around the concentration rises to the new set point but with a sustained oscillation of constant amplitude. This occurs only at the proportional band settings of (-350). The value then becomes our K_u. From the figure, the period of the sustained cycling is taken as the average time of two successive crests. To do this, the graph of Fig. 6.7 is expanded using the left top most icon to drag and enclose an expanded region with the help of the mouse to obtain Fig. 6.8. The time interval for a complete cycle was found to 94 seconds. Therefore P_u was assigned the value of 94 seconds. Using the equations of Table 6.2, the values for the controller parameters were then calculated and allocated to the PID block parameter. The calculated controller parameter values for the two methods are shown on Table 6.3.

Table 6.3 Estimated Parameter values for a PID controller

	Cohen-Coon Method	Ziegler-Nichols Method
t_d	10	
τ	61.62	
k_p	-0.018	
k_u	-	-350
P_u	-	94
k_c	-469.15	-206
τ_I	23.06	47
τ_D	3.532	11.75

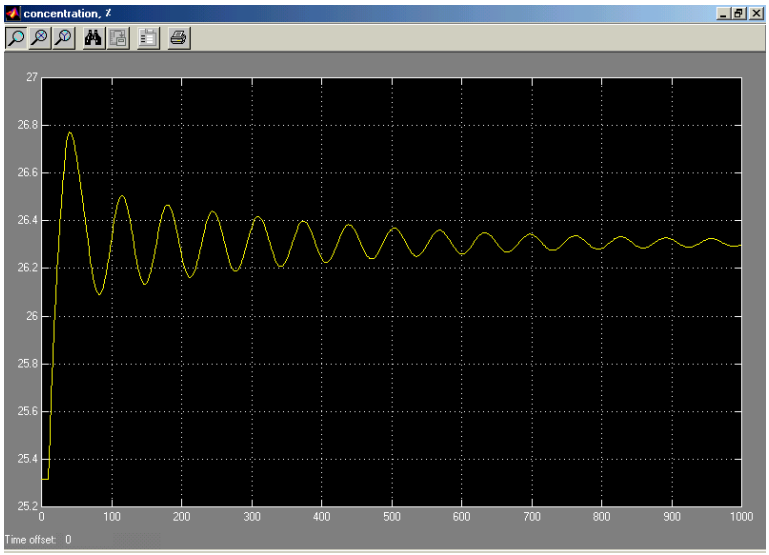


Fig. 6.6 System responses to set point change at gain of –320

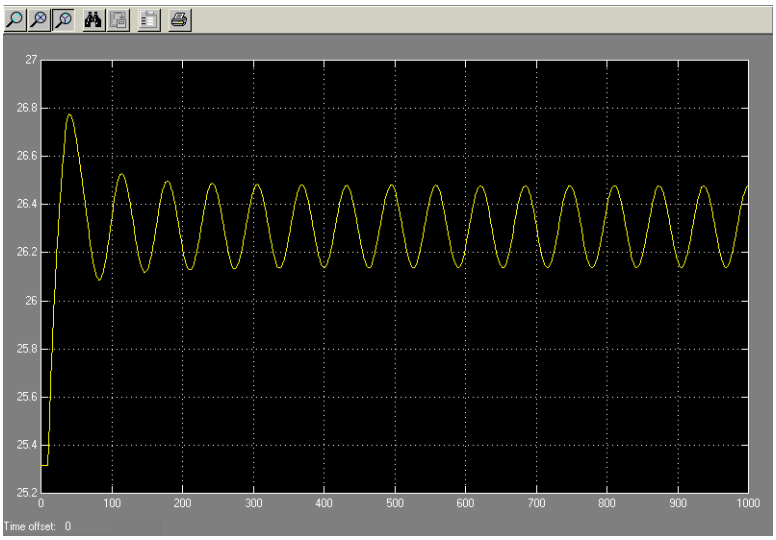


Fig. 6.7 System response to set point change at gain of –350

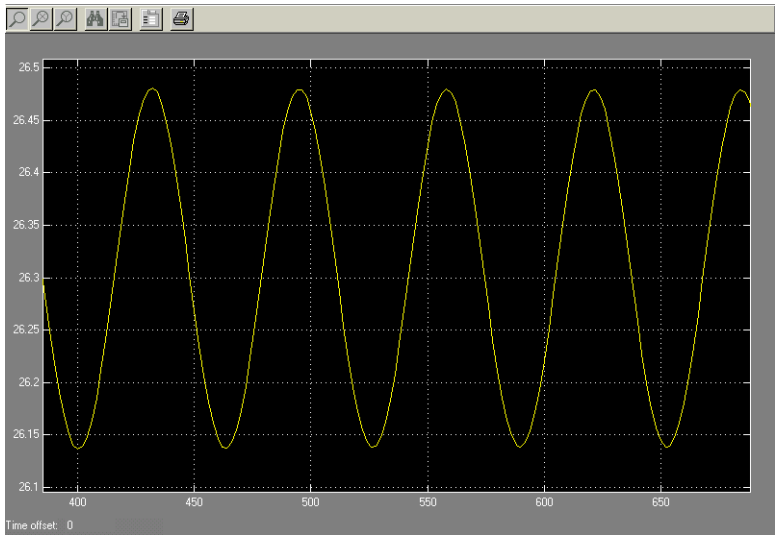


Fig. 6.8 Expanded Closed-loop response of concentration to set-point change

For the sake of rigid response, the controller parameters are re-defined in the software. The gain, k_c, is maintained as is, but the reset time is defined as integral gain $k_I = k_c/\tau_I$ and the derivative time as derivative gain $k_D = k_c\tau_D$. The modified parameters are listed in table 6.4. This issue is also discussed in chapter 1 section 1.4.3. It should be noted that the negative process gain and consequently proportional gain is the result of the increase/decrease behavior of the output-input (concentration-Steam Pressure) pair. The error signal used in the PID controller is based on the negative gain concept, i.e. $e = y^{sp} - y$. Therefore, if one wishes not to use negative values for the PID settings, which is the common practice in industrial application, then the definition of the error signal should be reversed. This modification can be easily handled, but we prefer to overlook it because we are simply dealing with simulations.

Table 6.4 Parameters used for the PID controller

	Cohen-Coon Method	Ziegler-Nichols Method
k_c	-469.15	-206
τ_I	23.06	47
τ_D	3.532	11.75
k_c/τ_I	-4.99	-4.39
k_c/τ_D	-406.2	-2421

6.3 TESTING THE ZN AND CC PID SETTINGS THROUGH SIMULATION

Using the values obtained in Table 6.3, the response of the process to set point changes and disturbances was tested. To do this the parameters to be used are recalculated to satisfy the requirement of the software. The values of these parameters are as given in Table 6.4. If these parameters are used and the response is unsatisfactory (highly oscillatory), the rule of thumb is to divide k_c by two and recalculate the other parameters by this new definition for the software. This is repeated until a satisfactory response is achieved.

For the set point change, double clicking on the set point icon of Fig. 6.1. The set point Block Parameter similar to Fig. 6.4 was obtained. The set point value was stepped from 25.316 to 26.316 by typing the new value. Press the enter button to complete this operation.

Now double click the PID controller icon to get the PID Block Parameter similar to Fig. 6.5. Then using the curser to highlight the parameter spaces enter the values for Proportional gain, Reset time gain and Derivative time gain for the Cohen-Coon settings as given in the last three rows of Table 6.4. Press enter and run the simulation to produce Fig. 6.9. Note in the response shown in Fig. 6.9, the CC proportional gain is divided by two because the original value produced very oscillatory response. Repeat the same steps for the Ziegler-Nichols settings to produce Fig. 6.10.

For the same set point change, Fig. 6.9 shows the product concentration rising to as high as 26.81 %. Then after two decays it settled at a steady state value of 26.316 %, showing no offset at all. The figure gave a rise time of about 60 seconds and a response time of close to 500 seconds. Fig. 6.10 produced almost a similar response. There was an overshoot but no decay, with a similar rise time to that of Fig. 6.9 but a faster response time of about 300 seconds. Not only did both settings produce similar responses but also the final steady state concentration values of 26.316 % for Cohen-Coon settings and 26.315 % for the Ziegler-Nichols settings are in very good agreement.

Judging from the responses depicted in Fig. 6.9 and 6.10, the only conclusion we can draw is that both methods provided excellent guesses for the values of the controller parameters.

In feedback control systems, controllers do not only act to keep responses as close as possible the desired set point changes but they also try to eliminate the impact of load changes from achieving the desired set point changes. Having already looked at how effectively our controller settings can cope with set point change, will an incoming disturbance sway the system from achieving the desired set point?

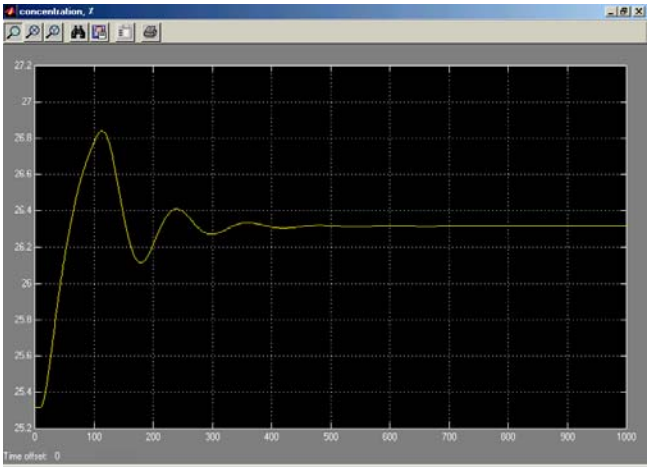


Fig. 6.9 Response of Evaporator Process to set point change using Cohen-Coon settings

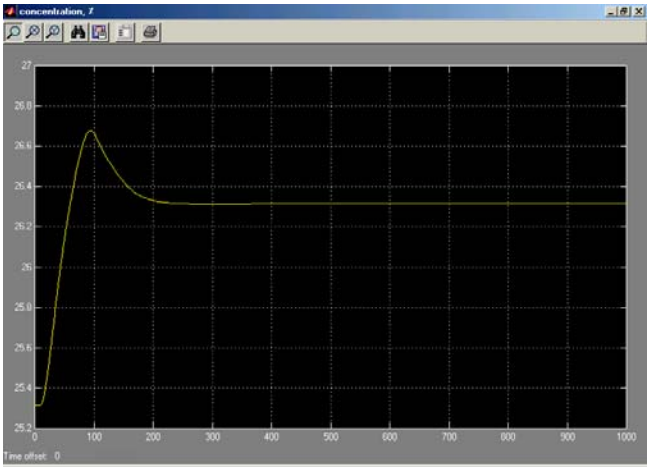


Fig. 6.10 Response of Evaporator Process to set point change using Ziegler-Nichol settings

To find out, the set point is maintained at its original value of 25.316 % and a disturbance introduced. Using Fig. 6.1, any of the input parameters (blue) can act as a disturbance. In this case the feed flow F_1 was chosen as the disturbance parameter. The flow was increased by a step change of 0.1 kg/min. This is done by double clicking on the F_1 icon to obtain a block parameter, from where the change is effected. Fig. 6.11 shows the open-loop response of the concentration to this disturbance. The open-loop response shows the concentration rising from the value of 25.316 % to a new steady state value of 26.36 % without limitation.

Fig. 6.12 and 6.13 show the controlled-responses to a 0.1 kg/min step change in F_1, the feed flow rate. Note how in both responses, the concentration rapidly rose to approach the value of the open loop response but was quickly brought back to the set point value of 25.316 %. Just like the set point response, the two graphs of the closed-loop system showed responses different from that of the open-loop response. Note the similarity between the responses of Fig. 6.12 and 6.13. The response controlled by the ZN settings settled faster with a response time of about 250 seconds as against 500 seconds for the CC settings. In both cases, the disturbance swiftly increased the concentration (as shown in the beginning of the two graphs) but the controllers acted immediately to bring back the concentration to the set point value of 25.316 %.

As concluded earlier, the controller setting parameters of both Cohen-Coon and Ziegler-Nichols are equally effective.

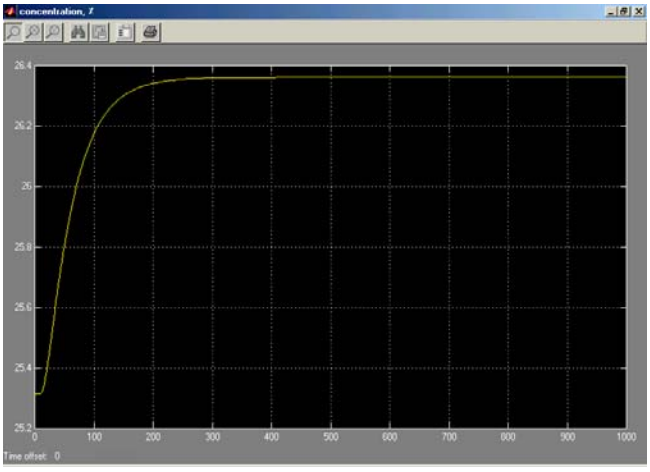


Fig. 6.11 Effect of Feed Flow disturbance on concentration in open-loop system

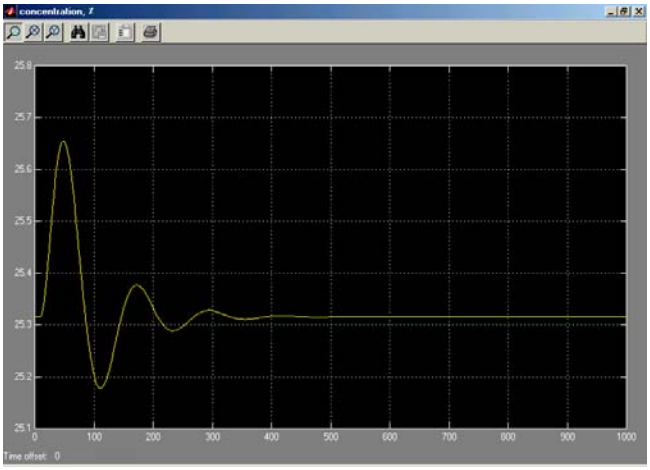


Fig. 6.12 Effect of feed flow disturbance on concentration in closed-loop system with Cohen-Coon settings

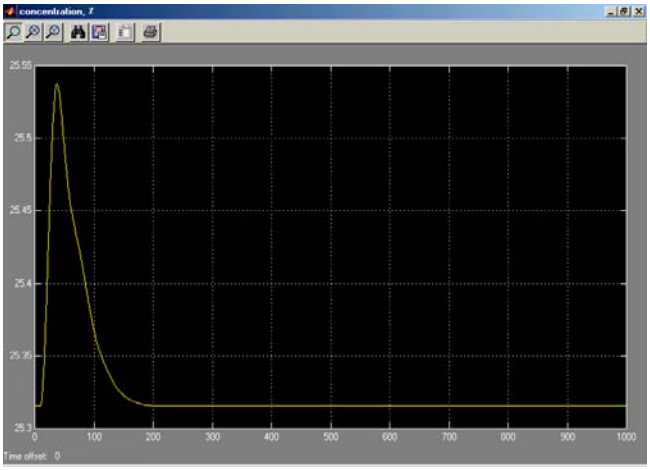


Fig. 6.13 Effect of feed flow disturbance on concentration in closed-loop system with Ziegler-Nichols settings

Remarks:

- The green box labeled “Go to menu” exists in the Simulink modules can be used to switch between the active Simulink module with the current active sub menu.
- The green box labeled “graphical tool” exists in the Simulink modules can be used to trigger the plotting forum. The usage of the plotting forum is discussed in Appendix A.
- The Saturation block in the Simulink modules is used to limit the input values between upper and lower constraints. The user can change the default values for the constraints by simply clicking the saturation block and insert the new values.

CHAPTER 7: MULTIVARIABLE SYSTEMS

7.1 MULTI CONTROL LOOPS

In chapter 6, the concept of single feedback control loop was discussed. Usually all chemical processes particularly the case studies presented in PCLAB have several process outputs. It is common practice that each input-output pair is linked through a single control loop. Therefore, the overall control system of a typical process or plant will contain multi single control loops. This idea is the scope of PCLAB exercise that will be discussed in this section.

Refer again to Fig. 1.16, one can simply choose the multiloop control option, By doing so, the following Simulink window pops up:

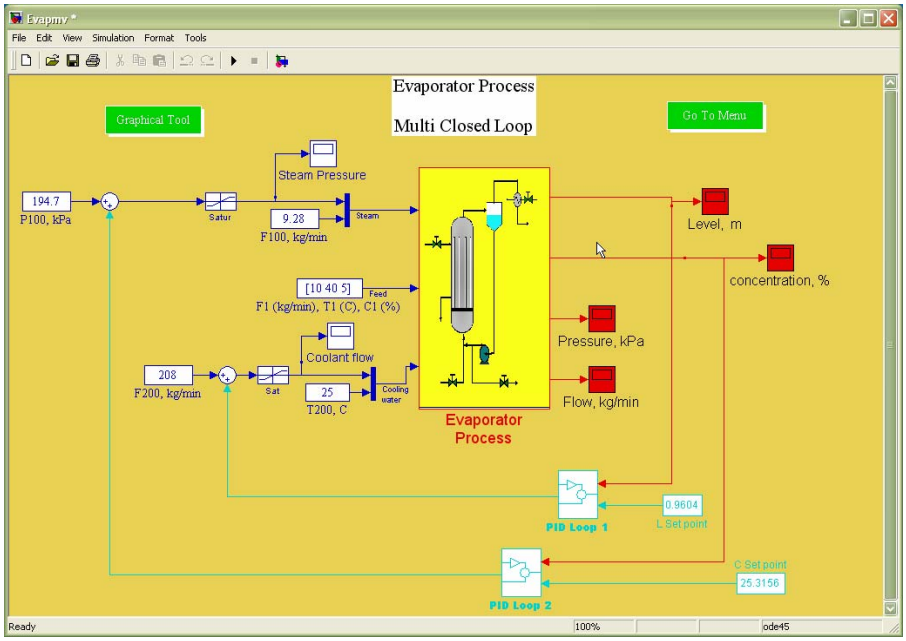


Fig. 7.1 Evaporator Process with two control loops

Fig. 7.1 shows two independent control loops. One loop pair the output concentration with the steam pressure and the other loop pairs the liquid level with the coolant flow rate. These two loops are chosen arbitrary simply for demonstration purposes. This means that the user can create as many control loops as the number of process outputs which is four in this case study. Moreover, the user can change the loop pairing, for example one can choose one of the available inputs (i.e., F100, F1, F200, T200) as the manipulated variable instead of P100 to control the product concentration. The modifications discussed above can be carried out by simple mouse operations.

However, it is suggested that only expert users can do so. The concept of selecting appropriate controlled variables, manipulated variables and pair them together is known as control structure design. This issue will be discussed later in this section.

7.1.1. Tuning of Multi Control Loops

As far as tuning is considered, the user can tune each loop independently using the procedure discussed in chapter 6. Following the Ziegler-Nichols procedure used in chapter 6 for each loop and with the aid of Table 6.2, the calculated PI settings are listed in table 7.1. Note that the ZN method is applied to each loop while the other loop is in open mode.

Table 7.1 Estimated Parameter values for a PID controller

	Loop 1	Loop 2
K_u	800	-350
P_u	50	94
K_c	360	-157.5
τ_I	41.6	78.3
K_I	8.6	-2

Next we examine the performance of the two control loops system. Let us introduce a disturbance in the form of a step change of magnitude 0.2% in the feed concentration, C_1. The process response in open-loop mode, i.e. with both controllers are disabled is shown in Fig. 7.2.

Fig. 7.2 Open-loop response to disturbance in feed concentration

It is obvious from Fig. 7.2 that the variation in the feed concentration affects only the output concentration. Now we repeat the simulation but with both control loops activated using the PI settings listed in table 7.1. The result is shown in Fig. 7.3.

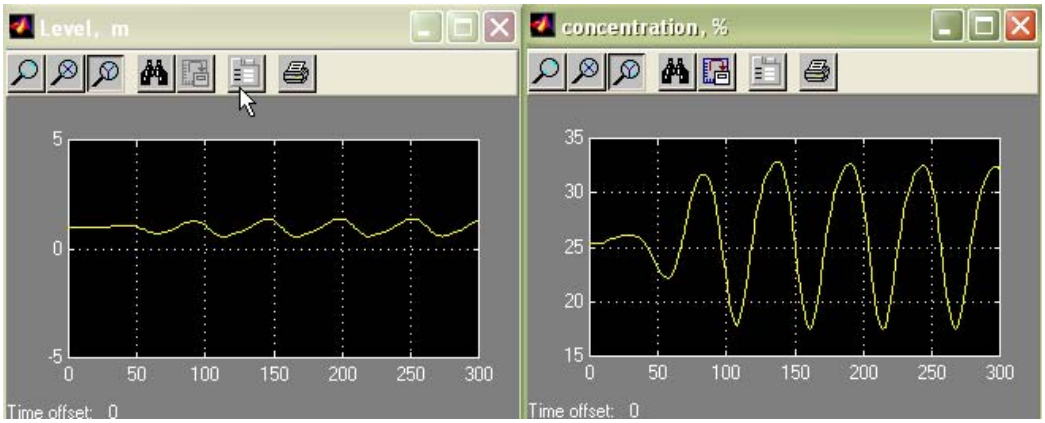


Fig. 7.3 Closed-loop response to disturbance in feed concentration

Fig. 7.3 illustrates that the response of the process is unstable despite the existence of the control system. This finding is not surprising because the loops were tuned independently ignoring their interaction. It is true that each loop will perform excellently when the other loop is open. However when both loops are active, the cross loop interaction is magnified causing poor and even unstable process dynamic behavior. A possible remedy is to fine tune both control loops. For example, if we cut down the PI settings of Table 6.1 by factor of 4 we obtain the results shown in Fig. 7.4.

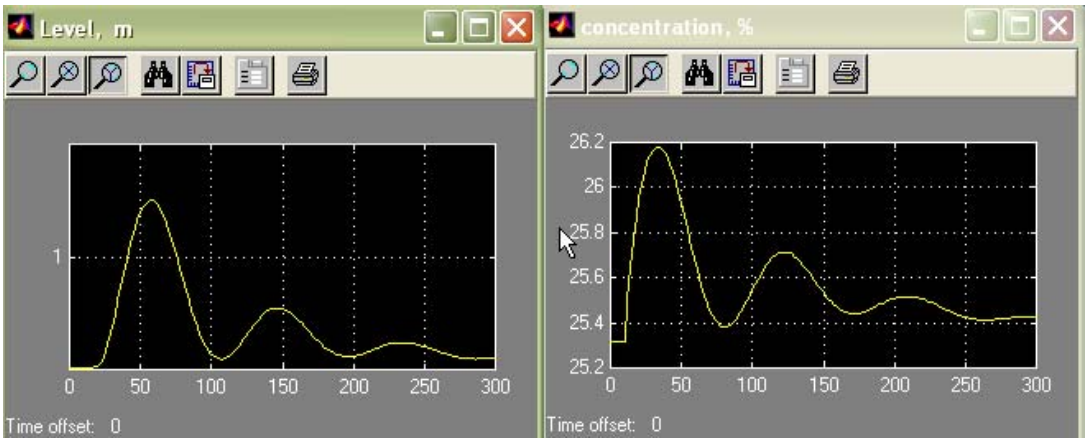


Fig. 7.4 Closed loop response to disturbance using improved PI settings.

There is no doubt that the resulting feedback response is much enhanced however it still suffers from oscillation. One can keep fine tuning the PI setting till desired performance is obtained.

Alternatively, one can use other tuning methods available for multi control loops system. Among these methods is the Biggest Log Modulus (BLT) proposed by Luyben [10]. The attractive feature of this procedure is simplicity since it includes designing only one parameter used as a de-tuning factor for all control loops. However, the resulted controller performance is conservative since the de-tuning factor is determined such that it provides tradeoff between stability and performance. Another design method is the Sequential Loop Closing (SLC) [11]. In this case, the loops are tuned individually but closed one after another so that interaction caused by closing a previous loop is accounted for during tuning the current loop. One drawback of such method is that interaction is taken care of in one direction only. This means that interaction brought by closing a current loop into all previous loops is not accounted for. Discussing these methods is out of scope. Nevertheless, users are encouraged to use the PCLAB platform to test these various procedures reported in literature and compare them.

7.1.2 Control Structure Design

As mentioned earlier control configuration which means selecting the controlled variables and their appropriate manipulated variables in multiloop's framework is an important design step for any successful control system. The control of a process with many variables can also be handled through multivariable control approach which requires advanced control strategies. The latter is out of the scope of the PCLAB.

There are several techniques available for control structure design [5]. The SSDSA method discussed in chapter 4 can also be used for control configuration. The SSDSA can be studied for each loop individually and consequently the appropriate structure can be determined. This approach however ignores the cross loop interaction. Alternatively, one can conduct the SSDSA method over all process variables and determine the best design structure. However, the method is based on steady state behavior only. Moreover, it is designed for a specific disturbance variable and hence may not necessarily work for other disturbances.

Another way to design a multivariable system is to use the concept of loop interaction. A Multivariable process is said to have interaction when process input affect more than one process output. The degree of interaction can be quantified by the so

called Relative Gain Array (RGA). Let K be the steady-state gain matrix of the process. Let $\boldsymbol{K}^{+}$ be the transpose of the inverse of the steady state gain matrix:

$$K^{+} = (K^{-1})^{T}$$

The elements of the RGA can be obtained as follows:

$$\lambda_{ij} = k_{ij} k_{ij}^{+}$$

The most important properties of the RGA are as follows:

1. The elements of the RGA across any row, or down any column sum up to 1.
2. λ_{ij} is dimensionless.
3. The value of λ_{ij} is a measure of the steady-state interaction:
    a. $\lambda_{ij} = 1$, implies that u_j affects y_i without interacting with the other control loops. *The system is completely decoupled*.
    b. $\lambda_{ij} = 0$, means that u_j has absolutely no effect on y_i. *Thus, y_i can not be controlled by u_j.*
    c. $0 < \lambda_{ij} < 1$, means an interaction exists, the smaller λ_{ij}, the larger the interaction is.
    d. $\lambda_{ij} < 0$, indicates strong opposite effect compared to its effect when other loops are open. *Such input/output pairing is potentially unstable and should be avoided*.

Based on the above results, RGA pairing rule is: *pair input and output variables that have positive RGA elements and closets to one.*

To compute the RGA, we can do the following steps:

Step 1: step the feed pressure, P100 by amount say 1 kPa with all control loops are disabled and record the reaction curve as shown in Fig. 7.5

Step 2: Using the response in Fig. 7.5 compute the steady state gain for all outputs as explained in chapter 5 and insert in the RGA matrix as the first column.

Step 3: Repeat step 1 and 2 for the other inputs, say F100, T200, F200. The final RGA matrix is:

$$K = \begin{bmatrix} -0.018 & -2.0 & 0.017 & -0.35 \\ -0.02 & 10.4 & 0.01 & -0.2 \\ 0.0037 & 10.7 & -0.08 & 1.4 \\ 0.0015 & -0.8 & 0.0067 & -0.12 \end{bmatrix}$$

The resulted RGA matrix is thus:

$$K = \begin{bmatrix} 0.3161 & 0.1164 & -4.19 & 4.76 \\ 0.0505 & 0.8704 & 0.332 & -0.253 \\ 0.096 & 0.2933 & 17.99 & -17.38 \\ 0.5374 & 0.28 & -13.133 & 13.87 \end{bmatrix}$$

The RGA matrix shows that the system is highly interacting. It is obvious that the third and fourth manipulated variables, i.e. F200 and T200, are collinear. We can see that they have the same effect on all output but at the opposite direction. This is expected because the coolant temperature and its flow has the same thermal effect but at the reverse directions. The first MV can be used to control either the liquid level (first output) or the output flow rate (fourth output). The second MV can be used to control product concentration (second output). Nevertheless, the control structure is not well defined. Another input, other than the coolant temperature, shall be examined.

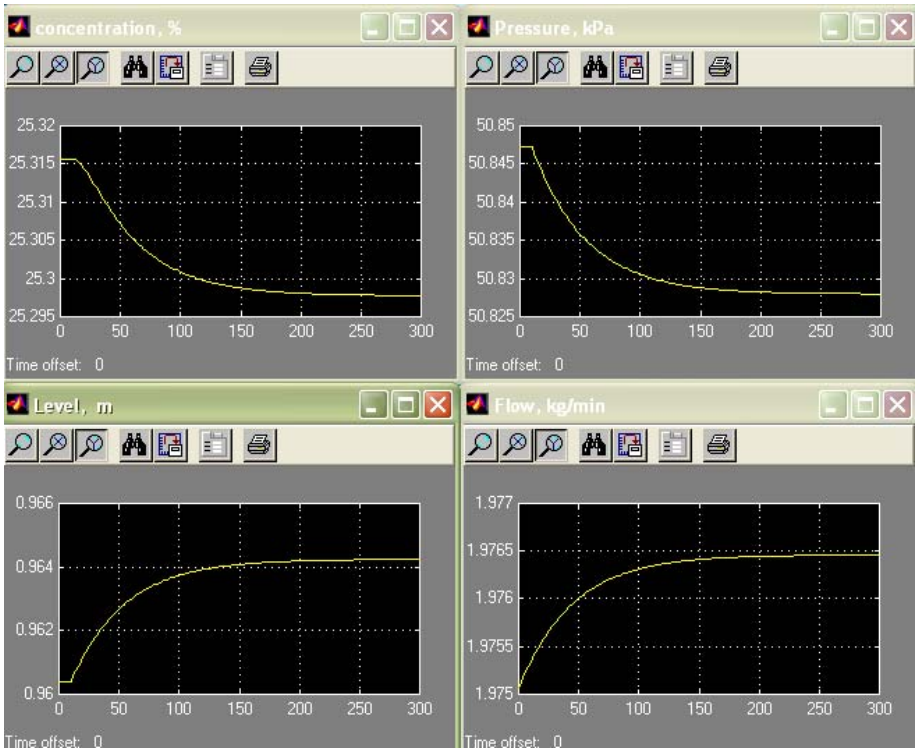


Fig. 7.5 Open loop response for a step change in the steam pressure

CHAPTER 8: ADVANCED CONTROL STRUCTURES

8.1 CASCADE CONTROL

Cascade control is one of the most successful configurations to improve the performances of the single-loop feedback control. It can provide more effective control by reducing both the maximum deviations and integral error for the disturbances response.

The concept of cascade control was introduced in chapter 3. The block diagram for cascade control is shown in Fig. 3.6 which consists of two control loops; the inner and the outer loop. The primary (also called outer, master) controller maintains the primary variable, say y_1 at its set point by adjusting the set point y^{sp}_2 of the secondary controller. The secondary (also called inner, slave) controller responds both to the output of the primary controller and to the secondary controlled variable y_2.

The key point in cascade control is the selection of secondary variable. Two guidelines must be observed:

- The secondary variable must indicate the occurrence of an important disturbance
- The secondary variable dynamics must be faster than that of the primary variable.

Let us study the cascade control through an example. In the main menu shown in Fig. 1.13, select “Multi-stage Flash Desalination” . This will bring the following submenu:

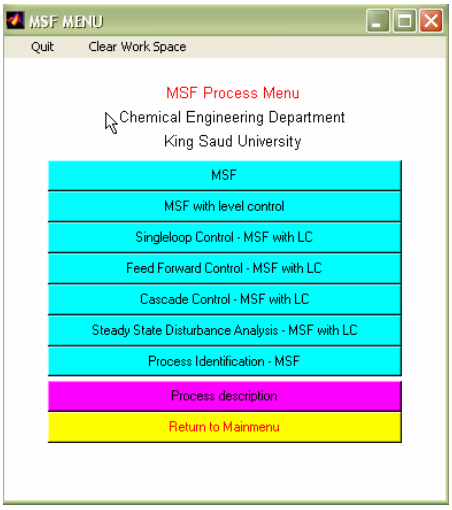


Fig. 8.1 The MSF Process Menu

Selecting “Singleloop control” in the MSF menu shown in Fig. 8.1 will bring up the following Simulink module:

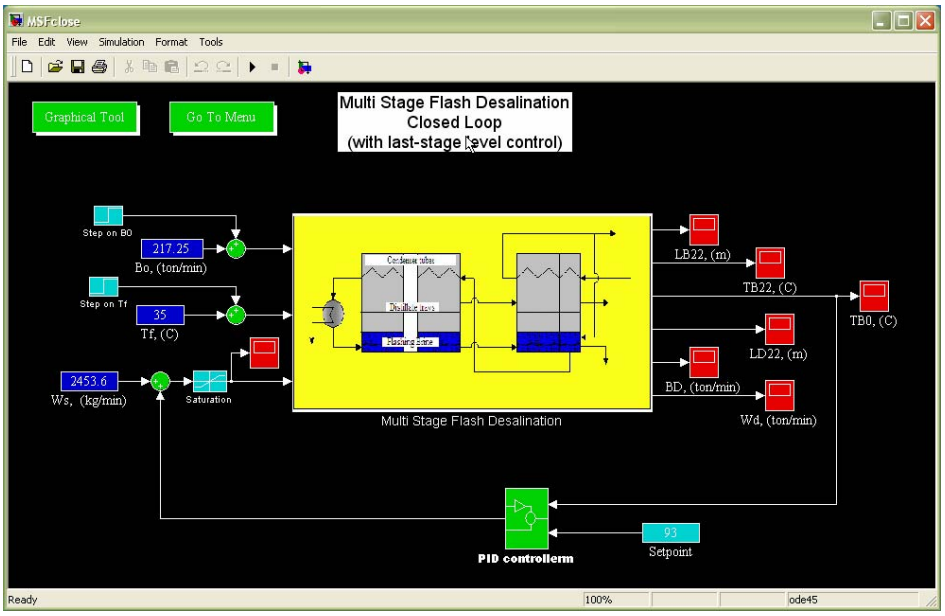


Fig. 8.2 Single loop control for MSF process

The module shown in Fig. 8.2 illustrates a control loop that regulates the top brine temperature, T_{B0} by manipulating the steam mass flow rate, W_s. The top brine temperature plays important role in the MSF operation as it affects the production rate, W_d significantly. Since the distillate product W_d, is the main output of the MSF, one may try to control the distillate product via manipulating the top brine temperature. However, the control objective can not be achieved directly. Therefore, one can work through a cascade loop. Selecting “cascade control” in Fig. 8.1 will activate the module shown in Fig. 8.3. The module shown in Fig. 8.3 contains a cascade control loop. The inner loop in that cascade structure is the T_{B0}-W_s control loop. This is the standard loop shown in Fig. 8.2. However, the set point of this loop is adjusted from an outer loop, which is the one designated for controlling the distillate product. Hence, the outer loop has W_d as its controlled variable and T_{B0}^{sp} as its manipulated variable.

Cascade control can use the standard PID feedback controllers in the two loops. The secondary loop must have the proportional mode but it does not require the reset mode. Integral model may be used in the secondary controller if it is desired to suppress completely the disturbance entering the primary or when the primary controller is not in operation (sensor not functioning or calibrated, etc). Derivative modes are not

advised in the secondary loop since the derivative action is designed to overcome some lag in the controller loop and if applied to set point changes may result in excessive valve motion and overshoot. The cascade controller is tuned in a sequential manner. The secondary controller is first tuned satisfactorily before the primary is tuned. Conventional tuning guidelines for PID, such those discussed in chapter 6, apply for both control loops.

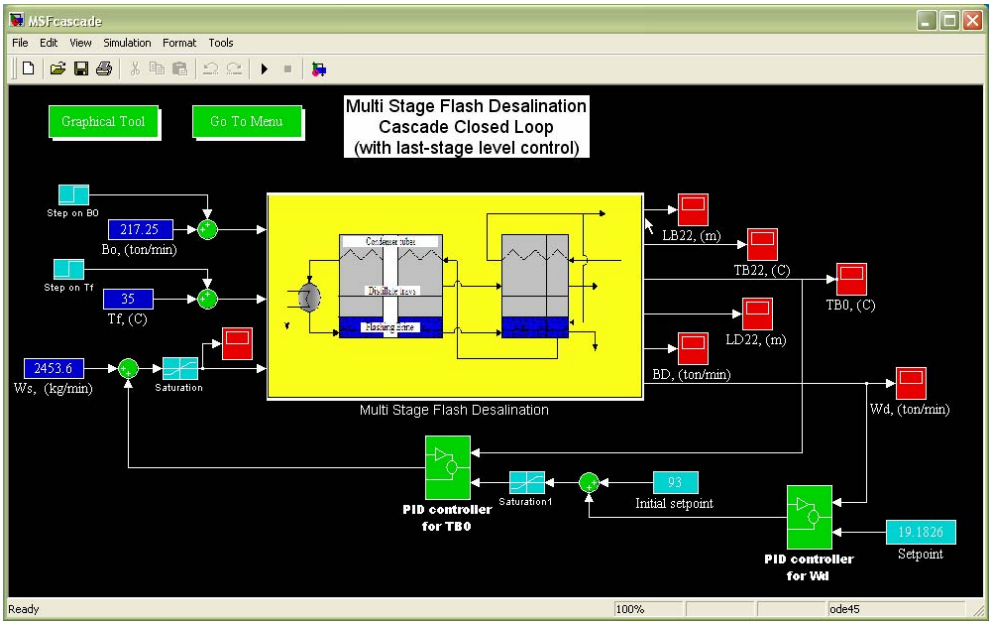


Fig. 8.3 Cascade Control Structure for MSF process

For satisfactory cascade control application, the inner loop must have faster dynamics. To check the speed of response to the inner loop, one can step change the steam mass flow rate with all loops disabled. As mentioned in chapter 6, disabling the control loop is achieved by settings all PID parameters to zero in the designated boxes, i.e. *PID controller for* T_{B0} and *PID controller for* W_d.

Introducing a step change in the steam mass flow rate result in the reaction curve depicted in Fig. 8.4. The curve indicates that the T_{B0}-W_s loop has a time constant of 20 minutes. It is hard to estimate the response speed of the W_d-T_{B0} loop because the top brine temperature can not be stepped independently. Nevertheless, the effect of Ws on T_{B0} is direct and its time constant controlled by the heat capacitance of the brine re-boiler. On the other hand, the effect of T_{B0} on W_d is transmitted through the 19 flash stages. This means that the time constant would be the accumulation of mass and heat

capacitance of all stages. According to this reasoning, it is evident that the speed of response for T_{B0}-W_s is faster than that for W_d-T_{B0}.

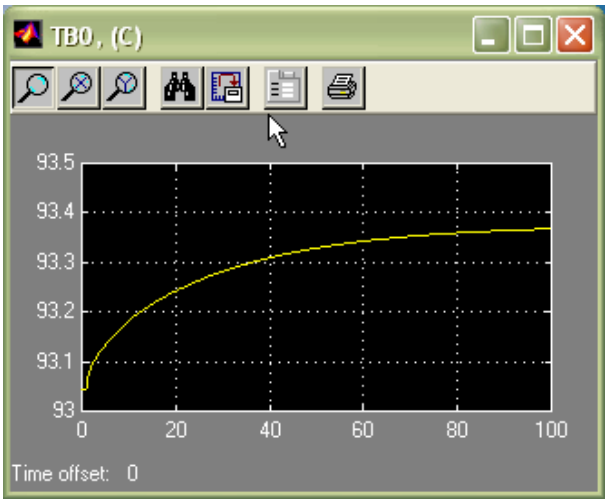


Fig. 8.4 The open loop response for T_{B0} to step change in W_s.

Fig. 8.5 demonstrates how the process responds to a disturbance of -5 ^{o}C in the feed temperature of the sea water without control.

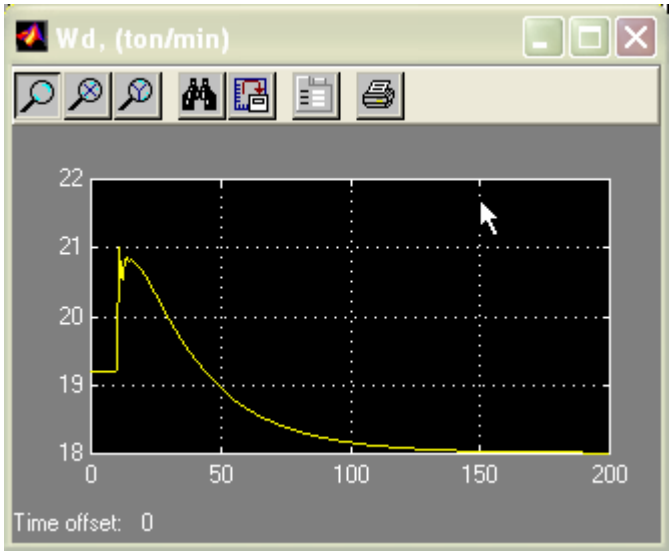


Fig 8.5 Product response to disturbance in feed temperature.

The result indicates a drop of more than one ton of distillate water. The situation demands a good control system. Fig. 8.6 illustrates how the process behaves when a single control loop like that shown in Fig. 8.2 is involved. The PI settings for that loop is $k_c = 50$, $k_I = 10$, $k_d = 0$. The feedback reaction in Fig. 8.6 demonstrates an excellent control of the top brine temperature but the product is left without control. Although the

product is increased, the feedback performance is considered poor from control point view.

Fig. 8.6 Product response to disturbance when single loop is involved

Now examine Fig. 8.7 which shows the process feedback performance when cascade control is installed. Here, the product flow rate, W_d is well maintained at its reference value. This is achieved by manipulating the top brine temperature to a new value of 88 ^oC. Note that maintaining the top brine temperature at its initial value is not necessary because it is not considered as a desired process output.

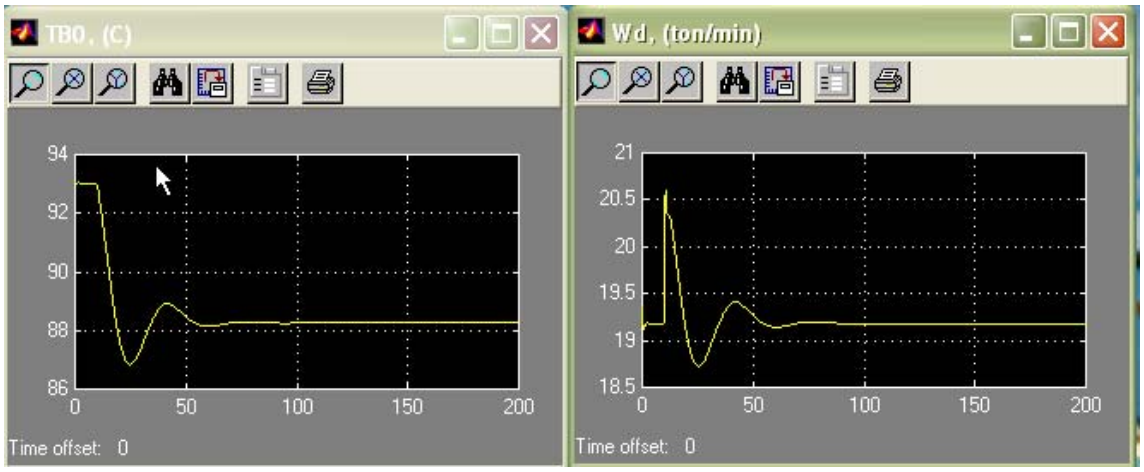


Fig. 8.7 Product response to disturbance when cascade control is involved.

The user can carry several control analysis on this cascade loop. For example, one can test the effect of tuning the inner and/or outer loop on the overall performance. This teaches how these two loops are interacting and hence care should be taken when tuning such loops. One can also test how the overall control loop performs when

disturbances enter the process. This teaches him how cascade loop in comparison to conventional loop can improve the feedback performance. Innovative users can create similar cascade control structure for the other case studies and examine their performance.

8.2 FEED FORWARD CONTROLLER (FFC)

Feed forward control attempts to enhance the performance of the single loop feedback control by making use of an additional measurement of process input as shown in Fig. 3.5. The implementation block diagram for FFC is shown in Fig. 3.6 and the design equation is given by (3.12) and (3.13). Feed-Forward should be used when feedback control does not provide satisfactory performance and a measured feed-forward variable is available. Comparison between the feedback and feed forward controllers is listed below.

Table 8.1 Comparison between Feedback and feed forward controllers

Advantages	disadvantages
Feed-Forward	
1. acts before the effect of a disturbance has been felt by the system 2. Is good for slow systems 3. It does not introduce instability in the closed-loop response	1. Requires identification of all possible disturbances and their measurement. 2. Can not cope with unmeasured disturbances. 3. Sensitive to process parameter variation. 4. Require good knowledge of the process model
Feedback	
1. Does not require identification of all possible disturbances and their measurement. 2. It is insensitive to modeling error. 3. It is insensitive to parameter changes.	1. It waits until the effect of disturbance is felt by the system. 2. It is unsatisfactory for slow systems. 3. It may create instability in the closed-loop response.

All case studies of the PCLAB contain feed forward design tutorial. For demonstration purposes we will examine the feed forward controller design for the evaporator process. Refer to the evaporator menu shown in Fig. 1.16 and select the feed forward controller. The following Simulink module shall appear:

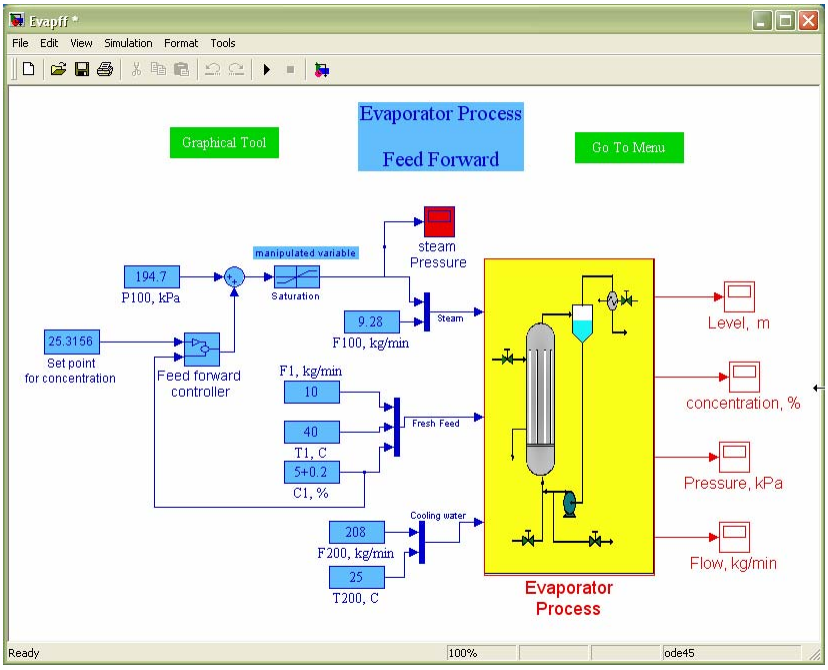


Fig. 8.8 feed forward Controller Structure for the Evaporator Process

Fig. 8.8 depicts a single FFC loop linking the disturbance C_1 (feed concentration) with the steam pressure. It should be noted this structure is not unique. The user can create, or replace it with, another FFC loop. According to the design equation discussed in chapter 3, the feed forward control block require the determination of the static gain and time constant for the manipulated variable and disturbance variable. These values can be inserted directly into the feed forward block as shown in Fig. 8.9. The dialog box for the feed forward controller shown in Fig. 8.9 can be activated by double clicking the "feedforward controller" icon.

In order to complete the FFC design, one needs to determine the value of the FFC parameters, i.e. k_p, k_d, τ_p and τ_d. these parameters can be computed using the reaction curve method discussed in chapter 3 and implemented in chapter 5. By step testing the product concentration to pre-defined changes in the steam pressure and feed concentration, the following were estimated:

Table 8.2 FFC tuning values

C-P100	$k_p = -0.018$; $\tau_p = 64$ s
C-C1	$k_d = 4.75$; $\tau_d = 22$ s

Next we insert these values in the feed forward parameter block and test the process for a step change of 0.2% in the feed concentration C_1. Fig. 8.10 illustrates the process response to the load change. Part (a) of the figure shows how the product concentration moves away from its nominal value when no controller of any type is involved. Part (b) demonstrates that the product concentration performance can be improved when a typical feedback controller is used. A typical PI controller is used with the controller gain and integral time magnitudes are set to the values found by Ziegler-Nichols method as listed in table 6.3. The last part of the figure depicts the process performance when FFC is implemented using the tuning values listed in Table 8.2. One can observe that the FFC response has less overshoot and faster settling time than those of the feedback controller. Nevertheless, FFC is expected to provide much superior performance than that of the feedback. The unsatisfactory performance of the FFC may contributed to the accuracy of the estimated parameters in Table 8.2.

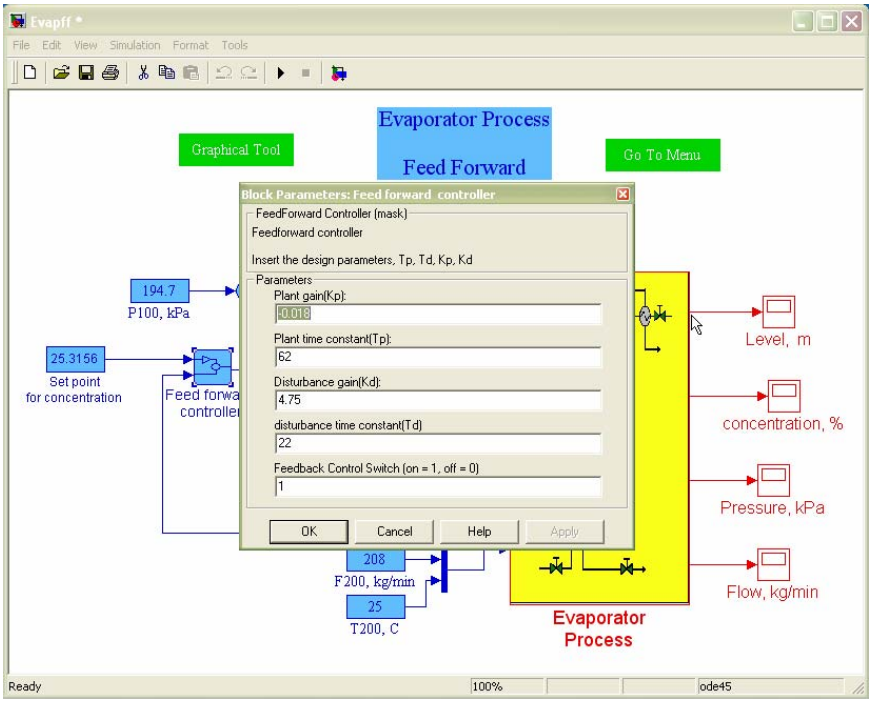
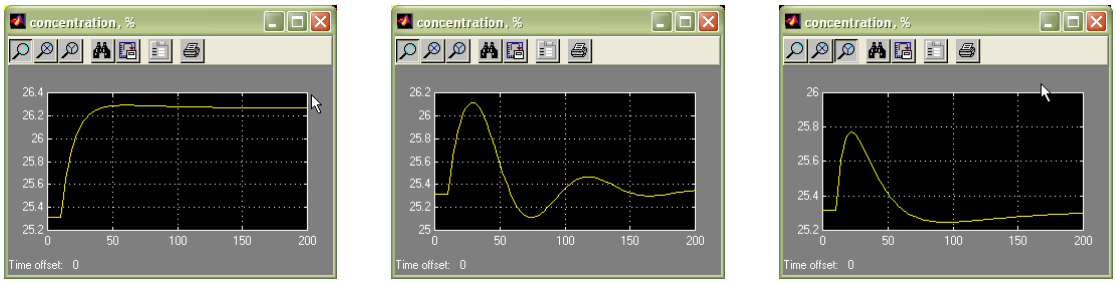


Fig. 8.9 FFC structure for Evaporator process showing the parameter dialog box.



(a) (b) (c)

Fig. 8.10 Process response to step change in the feed concentration

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APPENDIX A: SPECIAL FEATURES

A.1 PROCESS PARAMETERS

As discussed in chapter 2, each case study is comprised of differential equations with pre-specified parameter's values. Some key process parameters, which are believed to have important effect on the process behavior, are made accessible for modification. To access these parameters, one can simply double click on the process flow sheet shown inside the yellow box in any given Simulink module. Figure A.1 shows an example of accessing the process parameters for the evaporator process.

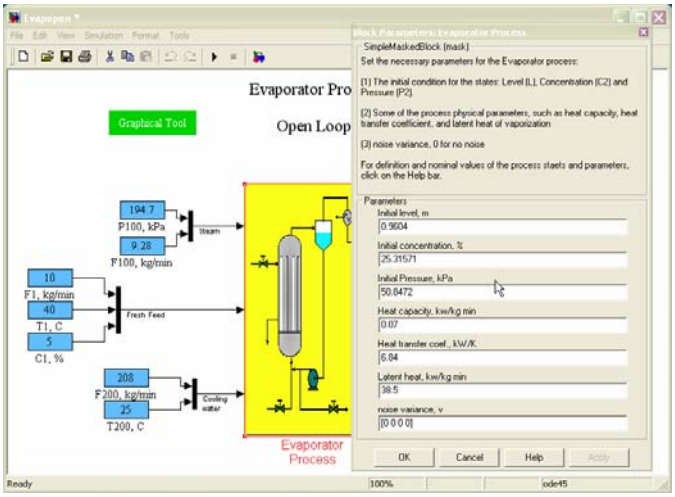


Fig. A.1 Accessing the process Parameter dialog Box

Specifically, the dialog box for the evaporator process contains that the initial value of the states, the heat capacity of the fluid, the overall heat transfer coefficient and the latent heat of the steam. The user can change the value of the designated process parameter directly. However, the user must run the simulation for open loop mode for enough long time to assure that the process reaches a new steady state. The new steady state values of the process states must be recorded in a separate sheet. The user should then access the process parameter box again and replace the initial value of the states with the newly obtained ones.

A.2 ADDING NOISE TO OUTPUT MEASUREMENT

All the simulation results shown in the manual depicted very smooth curves. Industrial practice indicates that output signals are always corrupted with noise. To add a realistic touch to the PCLAB simulations, a feature that allow the addition of measurement noise is adopted in this version of the software. As mentioned in the previous section, double clicking the yellow box in any Simulink module will open the process parameter dialog box, i.e. Fig. A.1. At the bottom of the parameter list appears the noise variance. The number of the noise variance equals the number of outputs. The order of the noise variance follows the order of the outputs appearance in the Simulink module. The user can specify the value of the noise *variance* which is supposed to be trivial compared to the variance of the measured output. Usually, noise is a random variable generated from mechanical and electrical sources. Since we are dealing with pure simulation, the noise is incorporated artificially to the data. Fig. A.2 shows the response of the product concentration in the evaporator process to 10kPa step change in the steam pressure. The left hand side plot depicts the response when no noise is preset while the right hand side plot shows the response with the measurement is corrupted by noise whose variance is 0.001.

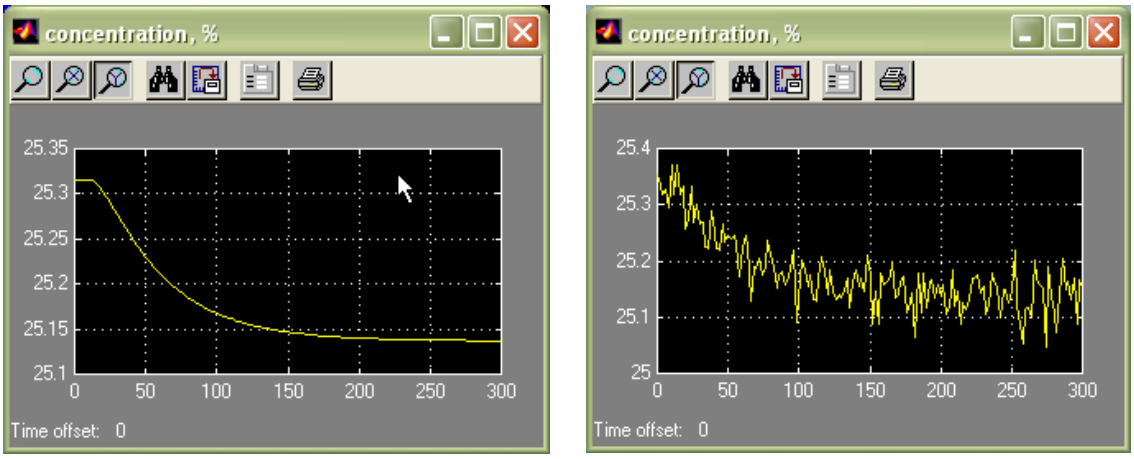


Fig. A.2 concentration response to step change in steam pressure

A.3 SIMULATION PARAMETER

All Simulink modules are simulated by pressing the start button as mentioned throughout the manual. In this case, the module will be simulated using default simulation parameters. To alter the simulation parameter, one can simply select *parameters* from the *simulation* menu as shown in Fig. A.3. This concept is also discussed in chapter 1 section 1.4.2. The user can adjust the simulation time, see Fig. A4. For example, if the simulation length is not enough to capture the full transient response, then the Stop time should be increased. Moreover, in some cases, the simulation may crash because the dynamic equations of the process are stiff or hard to integrate. In this case, the user should try different ODE solver. The current solver as it appears in Fig. A4 is Dormand-Prince. Pointing at the solver type box will reveal the available ODE solvers in a drop-down menu fashion. The user can also adjust the relative and absolute tolerance. In general reducing the tolerance will make the ODE easy to solve, but the resolution of the simulation plot will degrade. The user is also allowed to change the type of the solver from variable-step to fixed step.

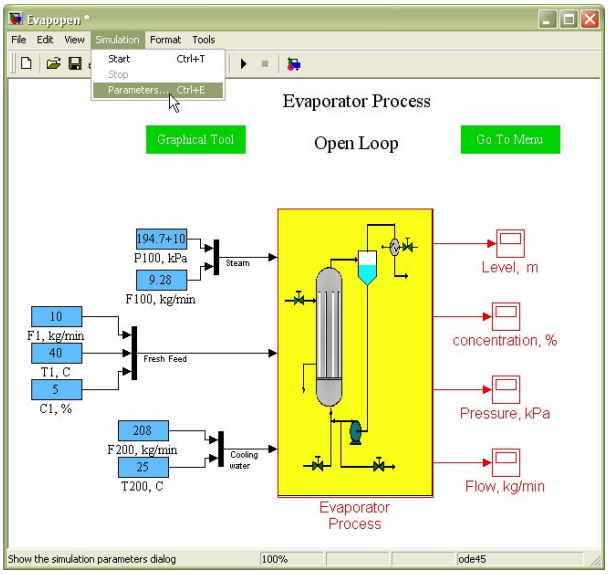


Fig. A3 Selecting the parameters option in the simulation menu item

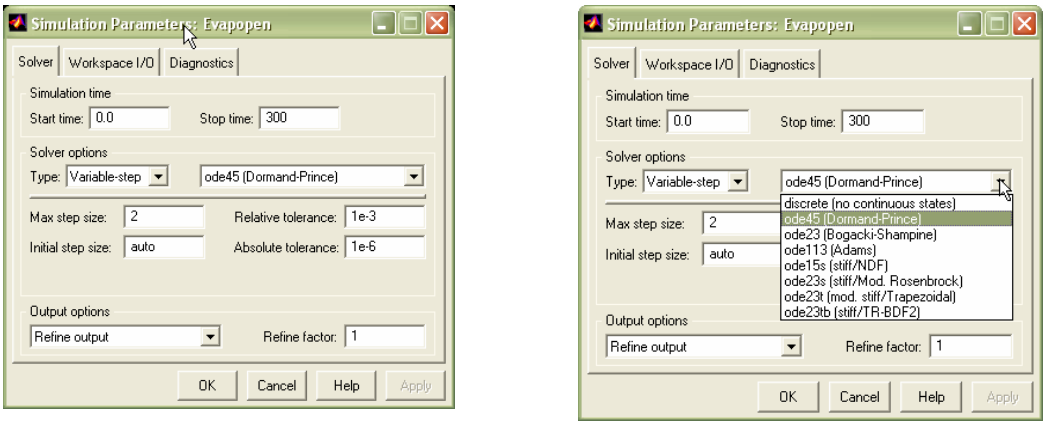
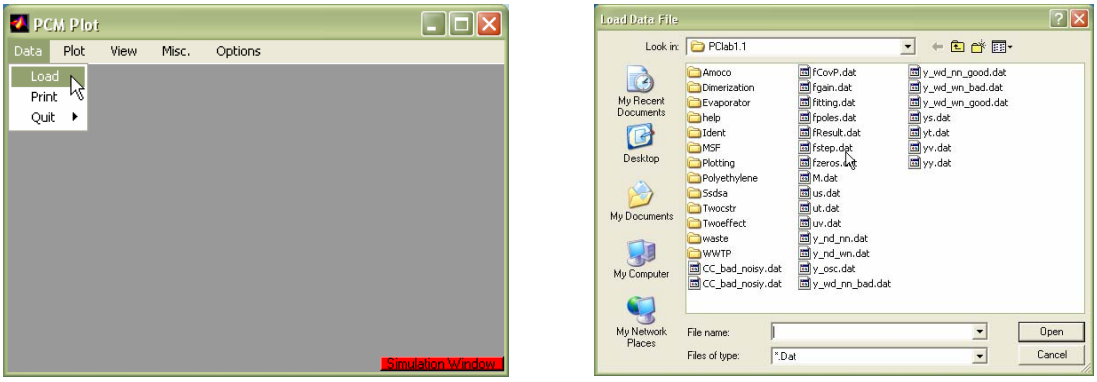


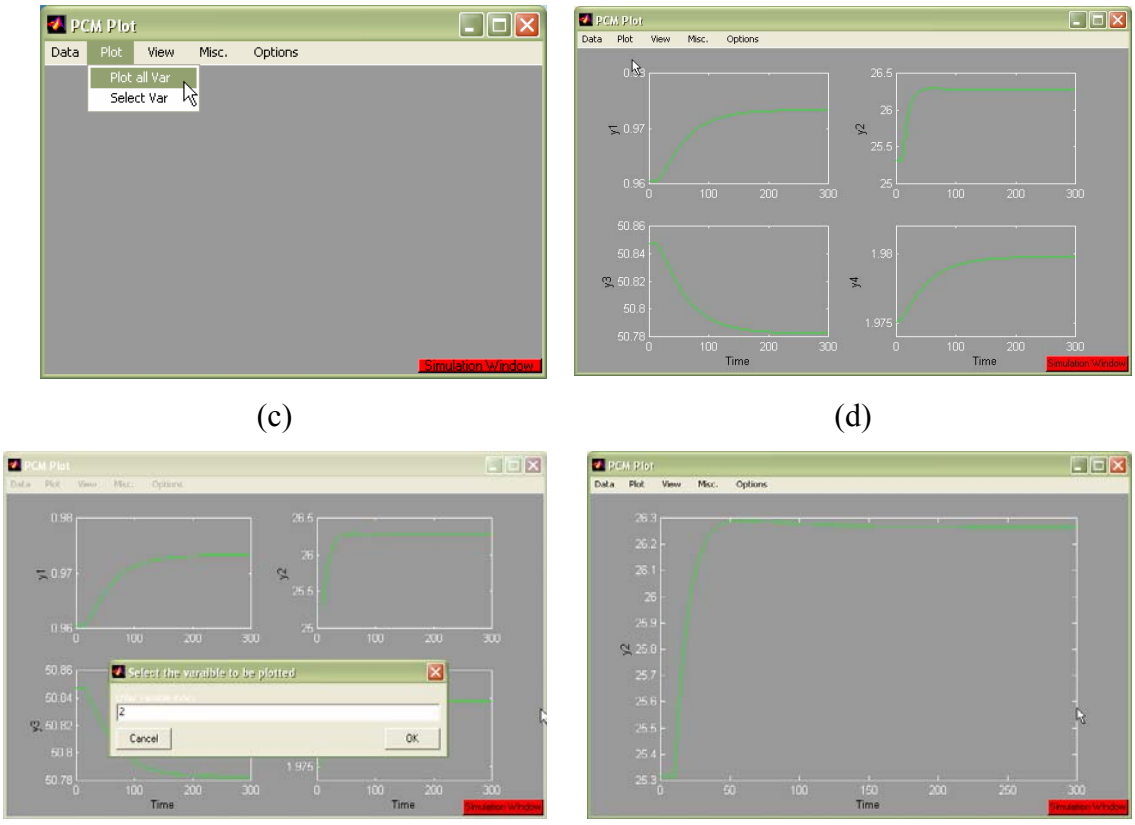
Fig. A4 the parameter adjustment dialog box

A.4 GRAPHICAL TOOL

All Simulink modules comprising the case studies allow for visualization of the transient response through the Scope box. However, the Scope shows only the result of the current simulation. Relying solely on the Scope capability, the user can not compare the result of different simulations in one screen. For this reason, we have added the graphical tool to allow the user to plot his results in a single screen. The graphical tool can be activated from any Simulink module by clicking the green box entitled "graphical tool". By doing so, the following window pop up:



(a) (b)



(c) (d)

(e) (f)

Fig. A5 Plotting Forum Options

Before starting the drawing, one should save his data containing the results of different simulation in a single matrix. The data matrix should have the sampling points in the first column. The following subsequent columns should contain the results for diverse simulations. After creation of the data file, the user starts uploading the data file by choosing *load* from the *Data* menu item, see Fig. A5a. By doing so, a new dialog box called "Load Data file" appears as in Fig. A5b pointing at the current PCLAB home directory. The user can select and load the desired data file or navigate through other directories for his desired data file. Next, the user can select "plot all var" as in Fig. A5c, which in turn draws all the data stored in the loaded file as shown in Fig. A5d. Alternatively, one can chose "plot selected var" which brings up the window shown in Fig. A5e. By inserting the number 2 to select the third column of the loaded data file, the plotting forum will be filled out by the response of the corresponding data signal.

If one tends to plot all results in a single box, then he should use the "plot selected var" option. Every time the user selects a specific variable index, it will be plotted in the

screen on top of the previous plot. Repeating this procedure will draw all variables in a single plotting box. The menu toolbar of the plotting forum has additional options such changing the axis scale, changing the foreground and background colors, adding labels, clearing the forum and a pointer. The latter can be used to point at a specific point of the plot and grip its local value.

APPENDIX B: TUTORIALS

B.1 SSDSA Open-loop tutorial: Exploring Steady State Behavior of an MSF Plant Case study

Objective: To study the steady state disturbance sensitivity analysis of an MSF in the open loop mode.

Note: In this tutorial, the MSF process is considered as an example. Similar procedure applies to other case studies.

Process description:

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2.

Here we analyze the static behavior of the MSF plant to variation in the process inputs (disturbances). This shows how the measured variable responds at steady state to a range of changes in the feed conditions. This analysis is essential for understanding the disturbance detrimental, severity and nonlinearity effect on the process. Specifically, the steady state behavior of the distillate product, W_d, Blow Down flow (B_D), Top brine temperature T_{B0}, last stage level and temperature (L_{B22}, T_{B22}) can be examined in the *simulink* module.

Launching the SSDSA simulation module

At the main menu of the PCCL, select the MSF case study. In the sub menu, select steady state disturbance analysis. A new simulink window, that shows the MSF in open-loop mode, pops up. One typical steady state operating condition for the input variables is shown in the corresponding boxes of the block diagram. Record these steady state values on the work sheet. To launch the SSDSA menu, simply click the *start* button.

Tutorial Procedure:

(1) Select one of the available disturbances by marking its checkbox.

(2) Select the open loop mode.

(3) Select appropriate step size and number of steps.

(4) Run the simulation by clicking the run button.

(5) Examine the generated plots.

(6) Discuss your results and make conclusions.

B.2 SSDSA Closed-loop tutorial: Exploring Steady State Behavior of an MSF Plant Case study

Objective: To study the steady state disturbance sensitivity analysis of an MSF in the closed loop mode.

Note: In this tutorial, the MSF process is considered as an example. Similar procedure applies to other case studies.

Process description:

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2.

Here we analyze the closed-loop static behavior of the MSF plant to variation in the process inputs (disturbances). This show how the measured variable responds at steady state to a range of changes in the feed conditions while certain inputs are used as manipulated variables, i.e. allowed to vary to compensate (counteract) for the disturbance effect. This analysis is essential for understanding the ability and efficiency of the chosen manipulated variable to control the process. Single loop and multiloops can be investigated.

Launching the SSDSA simulation module

At the main menu of the PCCL, select the MSF case study. In the sub menu, select steady state disturbance analysis. A new simulink window, that shows the MSF in open-loop mode, pops up. One typical steady state operating condition for the input variables is shown in the corresponding boxes of the block diagram. Record these steady state values on the work sheet. To launch the SSDSA menu, simply click the *start* button.

Tutorial Procedure:

1. Select one of the available disturbances by marking its checkbox.
2. Select the closed loop mode.
3. Select appropriate step size and number of steps.
4. Select appropriate manipulated and controlled variable pair.
5. Run the simulation by clicking the run button.
6. Examine the generated plots.
7. Discuss your results and make conclusions.

   Additional steps:
8. Repeat the above procedure for the same controlled variable but with another manipulated variable.
9. Compare the results for the two different manipulated variables and make a conclusion.

10. The user can reexamine all possible manipulated variables for the same controlled variable and comes up with the best input-output pair.

    Additional steps:

11. The user can study the multivariable framework by selecting two controlled variable and two manipulated variable simultaneously and run the simulation.
12. The user shall examine the results, discuss them and draw conclusions.

B.3 Open-loop tutorial: Exploring Dynamics of an MSF Plant Case study

Objective: To study the open loop dynamic behavior of an MSF plant and to fit the step test data to a first order plus dead time model.

Note: In this tutorial, the MSF process is considered as an example. Similar procedure applies to other case studies.

Process description:

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2.

Here we analyze the dynamic behavior of the MSF plant to variation in the process inputs. This show how the measured variable responds in time to changes in the feed conditions. This analysis is essential for understanding the dynamic characteristic of the process and for controller design. Specifically, the trend of the distillate product, W_d, Blow Down flow (B_D), Top brine temperature T_{B0}, last stage level and temperature (L_{B22}, T_{B22}) can be examined in the *simulink* module.

Launching the open-loop simulation module

At the main menu of the PCCL, select the MSF case study. In the sub menu, select MSF or MSF with level control. A new simulink window, that shows the MSF in open-loop mode, pops up. One typical steady state operating condition for the input variables is shown in the corresponding boxes of the block diagram. Record these steady state values on the work sheet. To examine the steady state value of the outputs at this operating condition, simply click the *start* button. Watch the output trend by clicking the designated scope icon to monitor the measured process variables.

For clear view of the output trend, the plotting scale shown in the scope can be adjusted either by clicking the left-button of the mouse on the edge of the data curve or by clicking the auto-scale button in the toolbar menu.

Tutorial Procedure:

1. Step the steam flow, W_s by a specific amount, say ΔW_s, click the start button and watch the temperature responds until it reaches a new steady state value. Record the step in the steam flow and the new Top Brine Temperature steady state value on the work sheet.

   Note:
   To step a specific input, simply double click the step function box that corresponds to the desired input. A new dialog box called "Block parameters" appears at which you can change the numerical value of the step input by entering a non-zero value for the "final value". You can also set the time at which the step change starts. Pressing ok will close the dialog box and save the changes.

   Some of the case studies do not have a designated step block. In this case, one can still impose step changes by directly clicking the desired input block. As a result, a block parameter dialog box pops up allowing the user to alter the value of the specific input variable.

2. Compute the steady state gain, the dead time if applicable and the response time as described in the work sheet.

   To estimate the final (steady state) value of the process output from the trend shown in the scope, simply draw a square around the end of the simulation using the left-button of the mouse. This will expand (zoom-in) that specific region. To get more precise value for the steady state, point at the curve and click the left button. This will increase the resolution of the y-axis. Similarly, to better estimate $t_{63.2}$, expand the region at which $y_{63.2}$ occurs using the same technique just described. Pointing the mouse on the curve and clicking the left button once more, the resolution of y-axis and x-axis will increases allowing more accurate reading of $t_{63.2}$.

3. Repeat step 3 using exactly the same amount of step change, ΔW_s, but with negative value. Record your results and compute the dynamic parameters as in the previous step.

4. The two steps in the steam flow were both the same in magnitude. Are the computed model parameters, i.e. K_p and τ_p, the same for these two steps? How/why are they different? Does the process dynamic behave linearly?

5. Repeat steps 3 to 8 for the other process inputs, i.e. B_0 (recycle flow rate), T_f (seawater feed temperature).

6. Based on the above analysis can you conclude upon which process input has the most effect on the process temperature? Can you determine which process input has the slowest impact on the process dynamics?

7. To further assess the linearity of the process dynamics, carry out two different but proportional step changes with specific ratio. Examine the resulted static gains and time constant. Are the calculated process parameters proportional to each others with the same given ratio?

Ideas

Innovative user can test the process dynamics for input functions other than step changes. The user can utilize the simulink library to includes, pulse, sine, ramp, etc signals.

Work sheet (*Effect of W_s on T_{B0}*)

Initial steady state:

T_{fss} =	W_{sss} =	B_{0ss} =	T_{B0ss} =

Positive step change	Negative step change
ΔW_s = New final value of T_{B0} = $\Delta u = \Delta W_s$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =	ΔW_s = New final value of T_{B0} = $\Delta u = \Delta W_s$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =
t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =	t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =
t_{step} = $\theta = t_{step} - t_{start}$ =	t_{step} = $\theta = t_{step} - t_{start}$ =

Work sheet (Effect of T_f on T_{B0})

Initial steady state:

T_{fss} =	W_{sss} =	B_{0ss} =	T_{B0ss} =

Positive step change	Negative step change
ΔT_f = New final value of T_{B0} = $\Delta u = \Delta T_f$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =	ΔT_f = New final value of T_{B0} = $\Delta u = \Delta T_f$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =
t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =	t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =
t_{step} = $\theta = t_{step} - t_{start}$ =	t_{step} = $\theta = t_{step} - t_{start}$ =

Work sheet (Effect of B_0 on T_{B0})

Initial steady state:

T_{fss} =	W_{sss} =	B_{0ss} =	T_{B0ss} =

Positive step change	Negative step change
ΔB_0 = New final value of T_{B0} = $\Delta u = \Delta B_0$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =	ΔB_0 = New final value of T_{B0} = $\Delta u = \Delta B_0$ $\Delta y = T_{B0} - T_{B0ss}$ $K_p = \Delta y/\Delta u$ =
t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =	t_{start} = $T_{B063.2} = T_{B0ss} + 0.632(\Delta y)$ = $t_{63.2}$ = $\tau = t_{63.2} - t_{start}$ =
t_{step} = $\theta = t_{step} - t_{start}$ =	t_{step} = $\theta = t_{step} - t_{start}$ =

B.4 Single-Loop PI Control Tutorial (1): PI Control of an MSF plant

Objective: To design and test a PI controller of an MSF plant for set point change and disturbance rejection.

Note: In this tutorial, the MSF process is considered as an example. Similar procedure applies to other case studies.

Process description:

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2. Here we analyze the dynamic behavior of the MSF plant under a single PI control loop.

All controllers should be designed for a specific level of operation, which include specifying an expected or desired value for set point and typical or baseline for important process parameters. Here we consider controlling the top brine temperature of an MSF plant. The main possible manipulated variable is the steam feed rate to brine heater.

A standard Proportional-Integral control algorithm will be used. It is therefore, necessary to determine the controller parameters, i.e. k_c (controller gain) and τ_I (integral time). Next the performance of the controller can be examined through simulation for tracking different set point changes and to reject possible disturbances. To simulate real practice, the PI control is implemented in discrete time fashion with sampling time of 5 minutes.

There are several methods for determining the PI parameter values. One method is the reaction curve method. This method utilizes the process model parameters determined in the previous Open-loop Workshop tutorial. These values can be fitted in predefined tuning formula such as the Coon and Cohen formula. Record your result:

Coon and Cohen formula:

$$k_c = \frac{1}{k_p}\frac{\tau}{\theta}[0.9 + \theta/12\tau] \qquad \tau_I = \frac{\theta[30 + 3(\theta/\tau)]}{9 + 20(\theta/\tau)},$$

$k_c =$ $\tau_I =$

If no dead time is observed, approximate the settings by:

$$k_c = \frac{1}{k_p}$$
$$\tau_I = \tau$$

Tutorial Procedure

1. At the main menu of the PCCL, select the MSF case study. In the sub menu, select MSF with single-loop control. A new simulink window, that shows the MSF in

closed-loop mode, pops up. Make sure that the controller parameters are set at zero. When the controller parameters are zero, the process will operate in open loop mode, i.e. without a controller. Click the start button to simulate the process. View the process response at the scope box and record the initial set point of the top brine temperature:

$$T_{B0\ setpoint} =$$

If the simulation need adjustment follow the instruction under "fixing the simulation time" at the end of this handout.

2. Enter the value of the controller gain and time integral computed in step 1 above in the designated boxes. Note that the reciprocal of the time integral multiplied by the sampling time should be entered in the *K*I box. Enter the initial set point value in the "set point" box. Rerun the simulation by clicking the start button. Watch the closed-loop response in the scope box. The *output* response shall remain at the recorded value above.

Set point test:

3. Explore the controller performance for set point change. Enter a new set point of 94 ^{o}C in the set point box, run the simulation and observe the temperature closed-loop response.

   To change the set point value, simply double click the "set point" box. When a new dialog box appears, enter your desired value in the space called "constant value".

   One can also examine the time response of the other outputs and the steam flow rate. At this given values for k_c and τ_I, does the temperature reach its new set pint exactly? Note that the final value of the temperature should be higher than the set point by amount of 0.046. Is the response fast enough?

4. Try different larger values for k_c at fixed τ_I and re-examine the closed-loop response. Try different larger values for τ_I at fixed k_c and examine the closed-loop response. In which case, the response is faster?

Disturbance test:

5. Explore the controller performance for regulatory problem. Re-enter the temperature set point of step 2 (without the decimal digits) in the designated box. Click the step box of the feed temperature. Add +5 to the given value. Close the dialog box by clicking *ok*. This step change simulates a sudden increase in the feed temperature. Set the controller gain and time integral to zero. Simulate the process under disturbance without controller. Observe the temperature response.

6. Re-enter the PI controller parameter values found in step 1 above. Click the start button and watch the top brine temperature responds. Discuss how the controller improves the regulatory response.

7. Repeat step 7-8 using a disturbance of magnitude of -5 in the feed temperature.

In the case of disturbance rejection, does bringing the top brine temperature to its original set point, brings also the distillate product to its original set point? why not?

Sampling time test:

All the above simulation was carried out at a specific sampling time of 5 minute. The choice of the sampling time plays an important role in the controller performance. To examine such a role, change the simulation sampling time. Double click the "*time*" box and change the sampling time to, say 10. Double click the large MSF box. When a new *simulink* model appears, double click the "*signal generator*" box. Update the period to 10.

After closing all dialog boxes, the new sampling time is activated. Run the closed-loop simulation (steps 4-9) again and observe the effect of the new sampling time on the control performance.

Idea

Innovative user can test different control structure, i.e. different output-input pairing.

Fixing the simulation time:

If the process response is fast, i.e. the simulation time is larger than the apparent dynamic of the process, change the simulation time. On the toolbar menu, open the "simulation" and select "parameters". A dialog box for *simulation parameter* comes up. Change the value of "Stop time" to your desired value.

B.5 Single-Loop PI Control Tutorial (2): PI Control of an MSF plant

Objective: To build intuition about PI tuning by trial and error method and experience the advantage and drawback of such a method.

Process description:

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2. Here we analyze the dynamic behavior of the MSF plant under a single PI control loop.

In previous tutorial, the controller parameters, i.e. k_c (controller gain) and τ_I (integral time) were determined by the reaction curve method. The reaction curve method requires building information about the process model. The reaction curve method requires stepping the process. Sometimes, it takes long time for the process to settle to another steady state value. During which disturbances may also occur. Moreover, some processes may not exhibit regular dynamic similar to that of a first-order system.

For this reason, PI controller can be tuned by the continuous cycling method as described here. The PI controller is applied in discrete time fashion with sampling time of 5 minute.

Tutorial procedure

1. Use the same MSF module with single loop control. Set controller integral time to zero. Therefore, the integral action will be disabled. Select arbitrary value for k_c. Similarly, select an arbitrary value for the set point. Click the start button to simulate the process. View the top brine temperature response at the scope box.

2. Repeat the previous step repeatedly. Each time increase the value of k_c gradually. Observe the temperature closed-loop response at each new value of k_c. When you observe a continuous cycling of the temperature response with steady amplitude, record the following information:

   The ultimate controller gain, k_u =

   The ultimate gain is the controller gain at which the continuous cycling is observed.

   The ultimate period, T_u =

   The ultimate period is the period of oscillation for the obtained continuous cycling.

   Using these ultimate values, compute the PI parameters as follows:

   $$k_c = k_u/2.2 \qquad k_i = 1.2T_s/T_u$$

   Where T_s is the sampling time. For clear capturing of the ultimate period of oscillation, you may need to zoom-in onto specific portion of the temperature response shown in the scope box. To zoom-in, simply point on the temperature curve, press the left-button of the mouse, draw a rectangular around the region to be zoomed, release the mouse button. An expanded plot of the zoomed-in region

will appear in the scope, from which you can estimate the cycling period. Clicking the mouse once more will increase the axis resolution even more allowing for better estimation of the numerical value.

3. Enter the computed values of the controller settings in their designated boxes. Explore the controller performance for set point change. Enter a value in the set point box, run the simulation and observe the temperature closed-loop response. Repeat the simulation for other set point values.

4. Explore the controller performance for regulatory problem. Re-enter the temperature set point of step 2 (obtained in previous Workshop (Tut_msf2)) in the designated box. Click the step box of the feed temperature (T_f). When the dialog box pops up, enter 5 for step time, 0 for initial value, +5 for final value and 5 for sample time. Close the dialog box by clicking *ok*. This step change simulates a sudden increase in the temperature of the seawater feed stream. Set the controller gain and time integral to zero. Simulate the process under disturbance without controller. Observe the process response.

5. Re-enter the PI controller parameter values found in step 2 above. Click the start button and watch the temperature responds. Compare the results with those obtained in the previous workshop, i.e. when the PI controller was tuned using reaction curve method.

   The controller settings found in step 2 above are for set point change. Repeat the same procedure, but for disturbance rejection. To do so, set the set point at the initial value reported in the previous Workshop. Step the feed temperature as in step 5. Repeat step 2 till the ultimate values are obtained. Are the new ultimate values exactly the same as those found in step 2?

6. Repeat steps 3 to 6 using the new controller settings **only** if they were different than these computed previously.

7. What conclusion can you make about the control performance using the continuous cycling tuning method?

B.6 Multivariable PI Control of Polyethylene Reactor

Objective: To build intuition about designing multi-loop PI control and experience the effect of cross-loop interaction.

Process Description

For detail description of the process, its main variables and schematic of the flow sheet refer to chapter 2. Here we deal the control of at least two controlled variables. For the specific polyethylene example, the monomer and hydrogen concentrations are controlled by manipulating their corresponding flow rates.

Tutorial procedure

6. At the Polyethylene sub menu, select multi-loop control. A *simulink* module, showing the closed-loop process in multi-loop mode, pops up. The first control-loop that links the monomer concentration with the monomer feed flow rate. The second control-loop links the hydrogen concentration with the hydrogen feed flow rate. These two control loops are chosen for demonstration purposes. The user can alter the control-loop structure as desired.

7. Compute the controller parameter values for each control loops using the same methods used in the single-loop control tutorials.

8. Insert the computed values of the controller settings in their designated boxes.

9. Explore the controller performance for regulatory problem. To do so, select an appropriate disturbance from the available list. Change the numerical of the chosen disturbance. Simulate the process under disturbance with the two controller loops in service. Observe the output response.

10. Discuss the performance of the multi loops control system. Do both controlled variable return back to their set points rapidly and smoothly? Try to increase the controller gain or time integral of the second loop and rerun the closed-loop simulation. What do you observe? And Why?

11. Change the upper and lower limit of the feed flow rates. To do so, point the mouse on the PID subsystem, press the mouse right button and select "look under mask". When a new *dialog box* appears, double click the "saturation" box and change the upper and lower limit values. Close the dialog box.

12. Rerun the simulation for a larger value for the disturbance and observe the closed loop response of both outputs. Do you observe any sign of input saturation? If yes, do you observe any offset in the output responses?

13. Now increase the integral time for the second loop 10 times and rerun the simulation. Do you observe any improvement in the control performance? Is the performance of the other loop get affected by retuning? What do you conclude.

(*ideas*)

14. Look in the literature and/or books for tuning formula that apply for multi-loop PID control and try to implement it on this process.

15. In the above tests, the control pairing was predefined. It is more rigorous to pair the control loops based on Relative Gain Array (RGA). To do so, follow the following steps"

First compute the steady gain matrix as follows:

$$G = \begin{bmatrix} k_{pT,Tf} & k_{pT,Qc} \\ k_{pC,Tf} & k_{pC,Qc} \end{bmatrix}$$

Compute the inverse of the transpose of the gain matrix:

$$G_T = (G^T)^{-1}$$

Compute the RGA as follows:

$$RGA = \begin{bmatrix} g_{11} \bullet g_{T11} & g_{12} \bullet g_{T12} \\ g_{21} \bullet g_{T21} & g_{22} \bullet g_{T22} \end{bmatrix}$$

If the off-diagonal elements of the RGA are negative or positive close to 0.5, then the process is highly interactive. Therefore, decentralized control may not perform well.

Pair the controlled variable and manipulated variable so that the corresponding relative gain (row element) is positive and as close to one as possible. According to this analysis does the pairing suggested by the RGA in consistent with the pairing used in the above tests?